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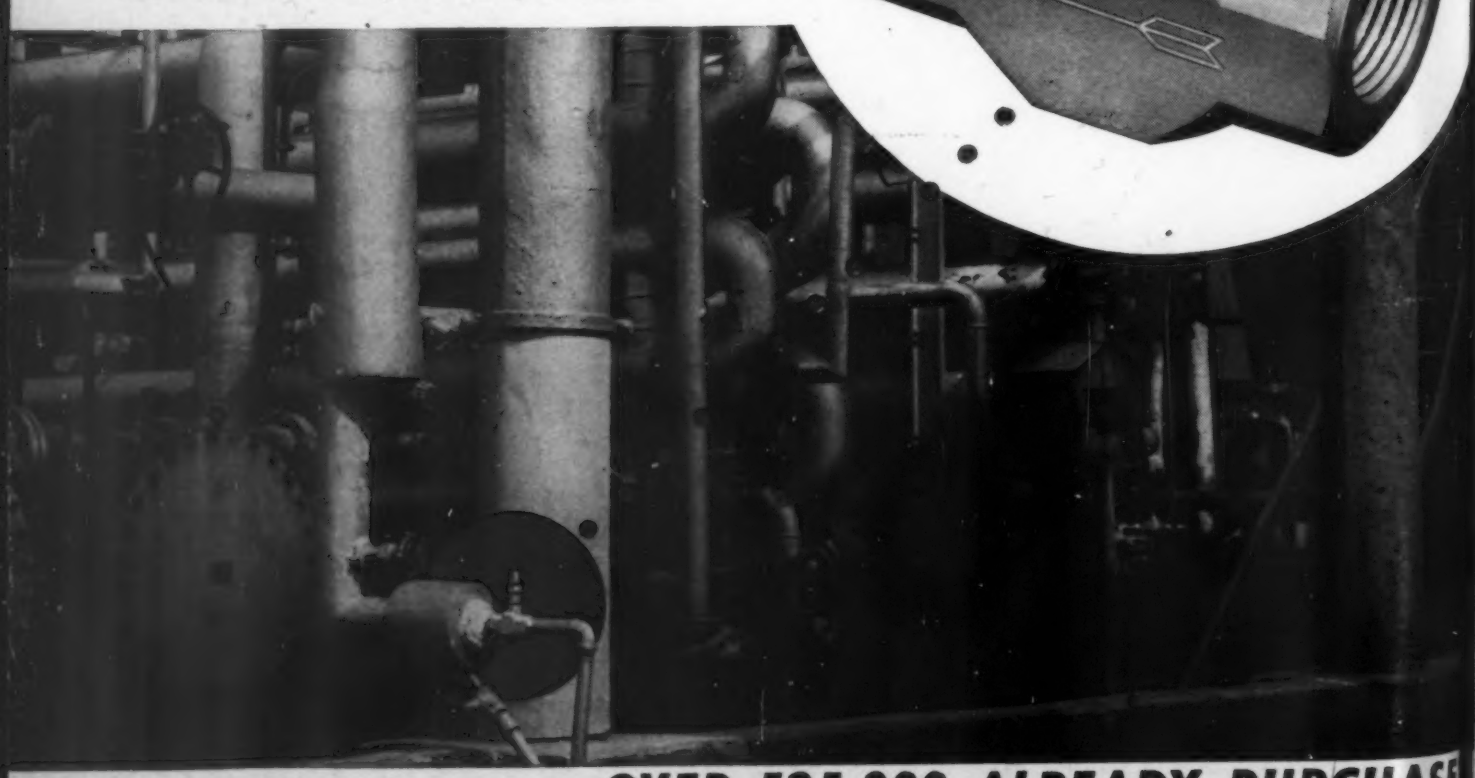

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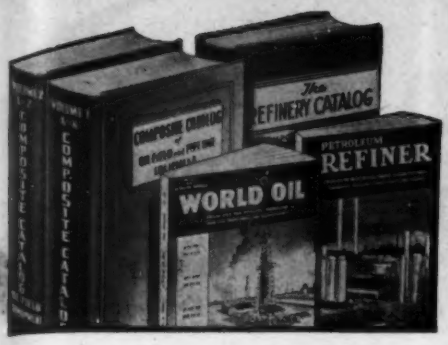

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for the refining-natural gasoline industry

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PETROLEUM REFINER

Specialized for the manufacturing division of the petroleum industry, refining, petro-chemical processing, cycling and natural gasoline operations.

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Now! Proportional-Reset

The Proportional-Reset Level Controller combines the basic Proportional Level Controller with automatic reset response. The addition of reset action eliminates the permanent deviation or "offset" of the level resulting from load changes which is the inherent characteristic of proportional controllers.

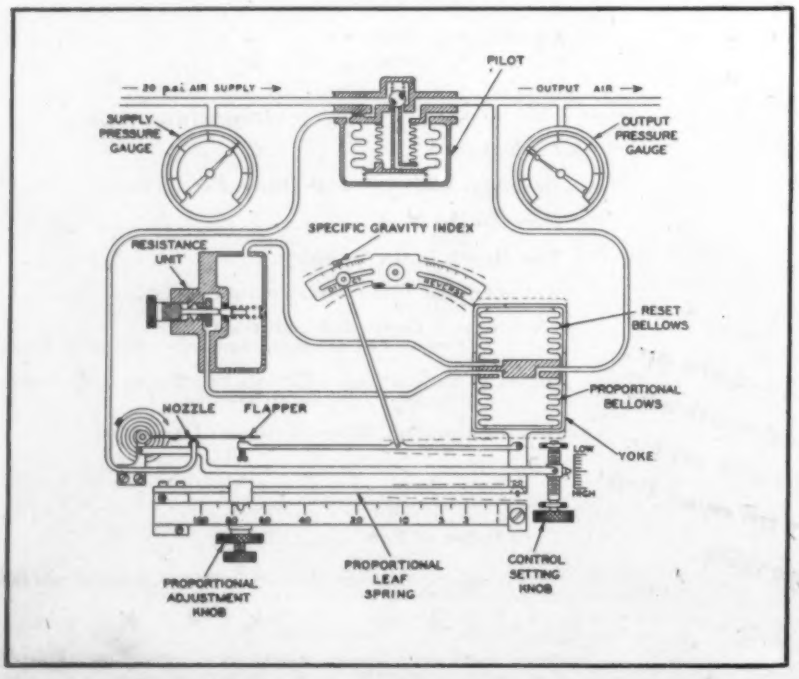
CONSTRUCTION

The Proportional-Reset Controller differs from the standard Proportional Controller in the addition of a reset bellows pneumatically connected to the proportional bellows through an adjustable resistance. The free ends of the opposing reset and proportional bellows are rigidly connected by means of a yoke. The force exerted on the proportional leaf (cantilever) spring is proportional to the algebraic sum of the pressures in the proportional and reset bellows. The proportional band adjustment is made by varying the effective length of the leaf spring.

The resistance unit consists primarily of a precision machined stainless steel tapered plug and tapered orifice which controls the flow of air between the proportional and reset bellows. The position of the tapered plug, which determines the reset rate, is set by rotating the adjustment knob. A spiral scale allows two full turns of the reset knob and is graduated in arbitrary units of 0.2 to 100 which are directly proportional to actual reset rate.

APPLICATION

The Proportional-Reset Level Controller is used for: (1) applications where constant level is required in the presence of large sustained load changes and (2) "averaging" level control. Standard ranges from 14 inches to 15 feet. Specific gravity ranges are 0.5 to 1.2 and 0.8 to 2.0. Available torque tube materials are inconel, Type 316 stainless steel and monel.



VOLUME 26, NO. 12 DECEMBER, 1947

PETROLEUM
REFINER

SPECIALIZED FOR THE OIL REFINING INDUSTRY, INCLUDING PETRO-CHEMICAL ENGINEERING, PETROLEUM SYNTHETICS AND NATURAL GASOLINE

A GULF PUBLISHING COMPANY PUBLICATION

THE LOOK BOX

THE Manufacturing Department of the Socony-Vacuum Oil Company has for the second time in two years achieved the best safety record of any large manufacturing department in the oil industry, according to the scoring method of the National Safety Congress, and has been awarded a trophy for their outstanding record. This record was achieved by an average of 8275 employees during a twelve month period. They worked 17,061,988 hours with 4314 days lost, an average of 2.50 days for each 10,000 hours worked. By National Safety Congress scoring, Socony-Vacuum had an accident rating of 4.78 compared with an average of 18.77 for the 17 major oil companies which participated in the contest.

Record Of Safety

High praise to Socony-Vacuum's manufacturing department! Its continued good showing is noteworthy, and serves to prove that a fine safety record it not a matter of chance but is the result of an endless effort on the part of an entire organization. Companies with less attractive safety records will do well to follow their example.

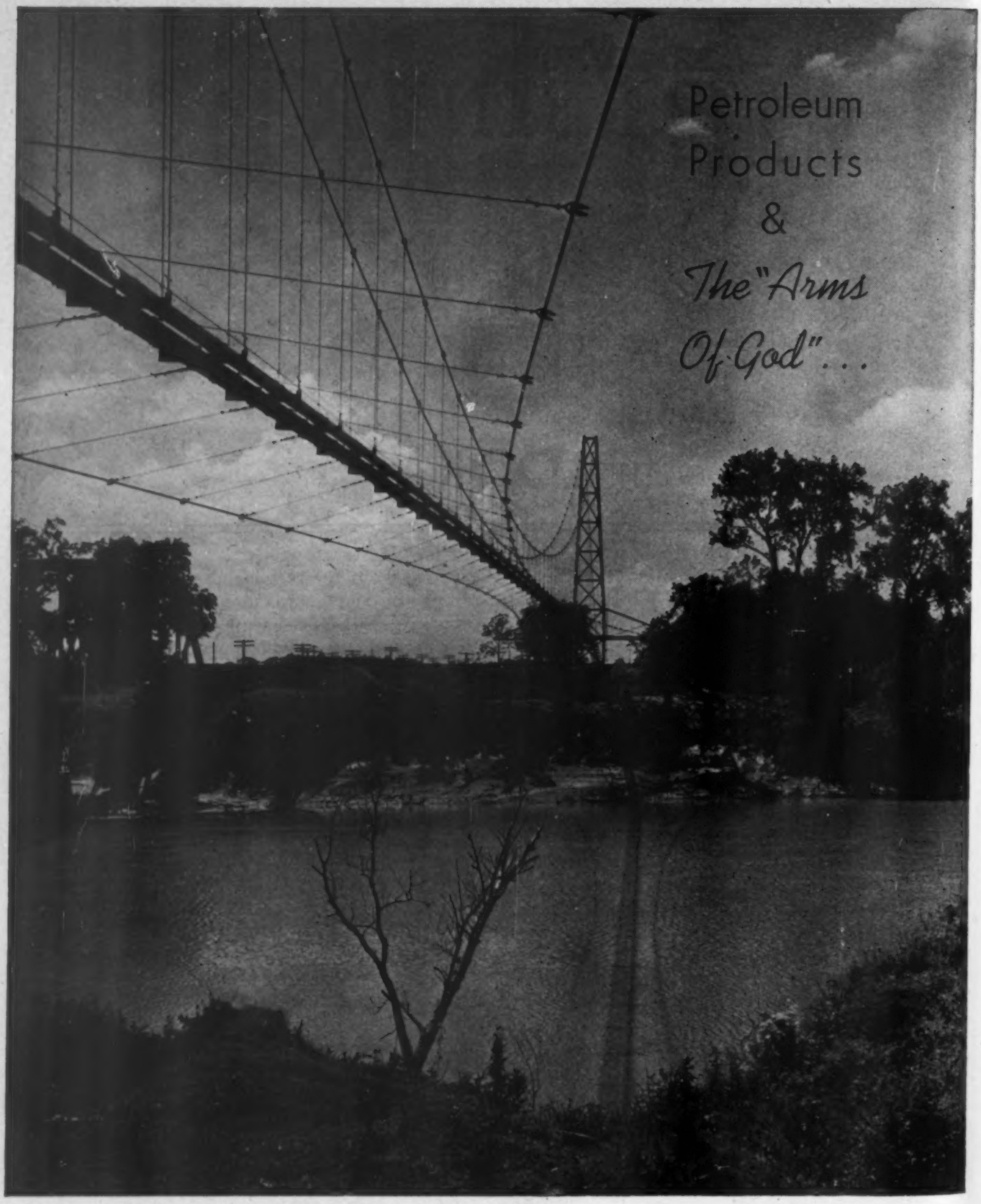
THE opinion has been expressed by many that the administration request for a return to general price control and rationing is motivated more by political considerations than by any firm belief that such a measure can be put through Congress. There is, however, some prospect that selective price control of a few basic items, including oil, coal, and grains, may be reenacted.

Oil Is Doing A Heroic Job

Insofar as oil is concerned, the difficulties being experienced by the petroleum industry in meeting record consumptive requirements, coupled with proposals for establishment of government controls to curb the inflationary spiral, could easily lead to the imposing of controls almost as severe as during the war. The clamor for controls may increase sharply during the next few months, especially if some parts of the country suffer because of insufficient fuel supplies. Even one percent of our population is nearly 1½ million people, and very many people either chilly, or out of work for lack of fuel oil, even for short periods, will expect remedial action. Rightly or wrongly, many of them will demand that the government do it or see that it is done. The pressure on the administration and on Congress may be great.

Referring to the administration proposal that it again be given authority to ration gasoline and control prices if need arises, A. W. Peake, president of Standard Oil Company (Indiana), recently said that such a plan under no circumstance would solve the oil supply problem, and that the oil industry can serve the public interest better working in its own way.

"No matter how well-meant they may be, government efforts to 'divide shortages fairly' will not produce a single extra gallon of oil," Peake declared. "They will only discourage the normal process of free competition in which 34,000 oil companies are now vying with each other to keep up with the greatest demand for oil products in all history. These companies are today giving America more gasoline and oil than the whole world was consuming ten years ago. The margin between demand and supply is small and could be erased by



Legend says Spaniards fleeing Indians in Texas long ago were saved when a sudden flood in a stream they had just crossed stopped the savages. Thankfully, they named the protecting river the "Arms of God," in Spanish, "Brazos del Dios," now shortened to Brazos. Modern Brazos floods seem heedless of whose path they block, but flow of petroleum products in the 8-inch line above will never the interrupted. Brown & Root, Inc. have bridged the Brazos near Sealy with a 625-foot (between towers) pipe line crossing of their design, fabrication and erection. Take your river problems to Brown & Root, Inc.

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only a little voluntary effort by consumers to hold down their needs while the industry works furiously to provide more facilities to enlarge the supply.

"Government intervention is the way to make the situation worse. Rationing and price controls would at once throw the whole distribution system into a tangle. The idea should not be entertained for a minute in peacetimes. The automatic mechanism of the market place will ration the products better than the government can. Price controls will also only discourage production. They are not needed anyway. Oil prices are not running wild. They are in fact lagging behind other commodities. The government index shows petroleum prices still below the 1926 figures while the price level for commodities in general was at last report at 153.6 percent of 1926.

"I hope that Congress will not be foolish enough to entrust to the administration the power to impost bureaucratic planning on an industry that is doing a herioc job of increasing its deliveries to consumers."

Matching an All-Time High

THE program of modernization and expansion within the oil industry is, like the demand for oil and oil products, at an all-time high.

This is pointed out succinctly in connection with the issuance of a booklet by the American Petroleum Institute which is urging fuel conservation in the home, and telling householders how to get the most from their fuel oil dollar.

The API feels that the material presented in the booklet will be of great interest at this time when the "unprecedented postwar demand for petroleum has created temporary problems in the prompt transportation and distribution of fuel to some sections of the United States."

The oil company's expansion program?

Well, during 1947-48 the bill will run more than $4 billion divided as follows: About $2 billion to drill new oil wells. More than $1 billion for new refinery facilities with a daily capacity of approximately 450,000 barrels of crude. Nearly a half billion dollars will build new pipe lines, extend old ones and construct other transportation equipment such as tank cars and barges.

Concurrently, we are reminded by W. J. Arnold, assistant vice president of The Pure Oil Company, the industry is engaged in a vast research program, some 10,000 scientists and research workers being employed by America's 34,000 oil companies.

Thus the industry is not only building, but, being a progressive segment of today's commercial picture, is doing its best to stay ahead through intensive scientific research.

$13 Billion Now Is Expansion Cost

IF YOU don't like API's figure on "more than $4 billion during 1947-48," given as the cost of the oil industry's 1947-48 program of modernization and expansion (see preceding item) then maybe you prefer the $13-billion mark which Dr. Joseph E. Pogue, vice president, Chase National Bank says the program will reach in providing "facilities with which to meet an expected 32 percent demand increase by 1951."

Writing in the *Journal of Commerce,* Dr. Pogue says that $7.5 billion (or 57 percent) of the total will be spent within the United States, 13 percent or $1.65 billion in the Middle East, 10 percent in Europe (excluding Russia), 9 percent in South America, 3 percent in the Far East and 8 percent in all other areas, (including Russia).

Previous to Dr. Pogue's new compilation, $4.5 billion was considered as the maximum figure for expansions in the United States during the present five year period. The $7.5 billion program in the United States will include exploration for and development of additional crude oil reserves, construction of crude oil and product pipe lines, increase of refining capacity by 400 thousand barrels per day and creation of additional marketing facilities, he discloses.

Pogue's world total is broken into the following categories: crude production, including exploration, drilling equipment etc., 53 percent ($7,000,000,000); refining facilities 23 percent, pipe lines 8 percent, marketing 11 percent, marine equipment 4 percent and all others 1 percent.

Nuclear Fuel Possibility

THE possibility of the discovery of a relatively harmless form of nuclear fuel resulting from the investigations of Drs. Nathan Sugarman and Anthony Turkevich, University of Chicago, of the deadly "ashes" left in atomic furnaces, has been disclosed. Such a fuel might do everything that coal, oil, or electricity does at a fraction of the cost.

Drs. Sugarman and Turkevich explained they had found that when an atom was split in two, various quantities of fission products are liberated, including some materials which are more lethal than others. The present sources of atomic energy release extremely lethal radiations and require massive installations of protective shielding. If they can find a source of energy liberating only comparatively mild radiations, the discovery would change the design of atomic energy plants, reducing the protective shielding which weighs 50 tons or more. Necessity of such massive shields has held back "development thinking" for actual plants, particularly those of a portable type.

Whether it's

Big Valves

like this - - - - - - - - - - - - -

or

Little Valves

like this - - - - -

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EAST CHICAGO, INDIANA

Lion Oil Company Starts Operation Of TCC Unit

F. L. EMERT, Plant Engineer

LION Oil Company, El Dorado, Ark., has recently completed and put in operation a new Thermofor catalytic cracking unit, which was licensed by Houdry Process Corporation and was designed and built by The Lummus Company. Designed and constructed at a cost of approximately $1.5 million, the unit required almost a year to complete. It was designed to process 4500 barrels per day of fresh gasoil with 900 barrels per day of recycle stock and has already demonstrated its ability to exceed both rated capacity and the guaranteed percentage yield of high octane gasoline. Though in operation for only a short time, the unit has been accepted by the management of Lion Oil Company.

General View of New Thermofor Catalytic Cracking Unit, Lion Oil Company's Refinery, El Dorado, Ark.

Process Design

The unit can be best described in three sections, namely: 1) feed preparation; 2) catalyst section; and 3) product separation.

The feed preparation section provides for a designed daily charge of 4500 barrels per day of gasoil charge mixed with 900 barrels per day of recycle stock. The fresh feed is preheated by the fractionator bottoms quench stock to 325° F. and mixed with 570° F. recycle stock with a resulting feed temperature of 370° F. The entire feed stock is first charged to a preheater furnace, then split and passed through two vaporizer furnaces in parallel. Prior to entering the vaporizer heaters, steam is added to the gasoil charge stock to aid in vaporization. The outlet temperature of the vaporizer heaters is designed for 830° F. The completely vaporized charge then enters a tar stripper where approximately 3.5 percent of the charge is removed as tar stripper bottoms, the overhead vapors entering a superheater where the vapors are superheated before entering the reactor. A portion of the tar stripper bottoms is used as recycle stock which is combined with the fresh feed. Another portion of the tar stripper bottoms is cooled and returned as quench for the tar stripper. The net tar stripper bottoms produced are removed from the system to fuel oil storage. The superheater outlet and reactor inlet temperatures are maintained at 880° F.

THE second small-scale Thermofor catalytic cracking unit recently to be put into operation and described herewith incorporates all the latest improvements.

The center of interest of the plant is the catalyst section, which consists of three units, a reactor, a kiln, and an elevator which carries the spent catalyst to the top of the kiln. The rest of the process is similar to the usual distillation unit.

The detailed description by the author will be of great interest to the refiners anticipating future investment and expansion of facilities along this line.

Catalyst Section

In the second, or catalyst section, the main reaction of the process occurs. This section is the heart of the process and contains the latest development in reactor design, that is, concurrent flow. The superheated feed enters the reactor at the top where it is contacted with the entering synthetic catalyst beads, which are maintained at a temperature of about 950° F. Catalytic cracking occurs as the two streams of vapors and catalyst contact each other and flow concurrently downward through the reactor. The catalyst and cracked vapors are separated in a disengaging section in the bottom of the reaction chamber. Residue oil on the catalyst is removed in the bottom section of the reactor by steam stripping. The oil vapors are passed to the fractionating section at a temperature of 880° F. where it is quenched with 370° F. bottoms quench stock; while the spent catalyst with its carbonaceous deposit is removed from the reactor by gravity flow into an elevator which carries it to a point where it will flow by gravity into the top of the kiln. The carbon laid down on the catalyst during the cracking re-

action is burned off in the kiln. The catalyst flows downward through the kiln which is equipped with nine burning and cooling zones. The combustion temperature is controlled by cooling coils and regulation of the combustion air rate. The heat generated in this operation is partially removed by circulating water through the cooling coils. This water is flashed into a drum where 225 psig steam is removed. Enough steam is generated in this operation to operate all the steam-driven equipment required for the entire unit. The regenerated catalyst is returned to the reactor hopper by means of the elevator and is again ready for reuse.

Mechanical Design

The third, or product separation section, is a conventional fractionating system. The oil vapor or "synthetic crude" from the reactor at 880° F. enters a 20-tray fractionator where a bottom product is removed as fuel oil, a side stream furnace oil fraction is removed and the overhead of gasoline and lighter gases are removed. A gasoline having an end point of 425° F. is produced. The gases removed from the system are compressed and the pentanes and heavier are condensed and stabilized as gasoline, the butanes and lighter are further processed in the vapor recovery unit as butanes and gases with a fraction going to poly plant for polymerization to high octane gasoline.

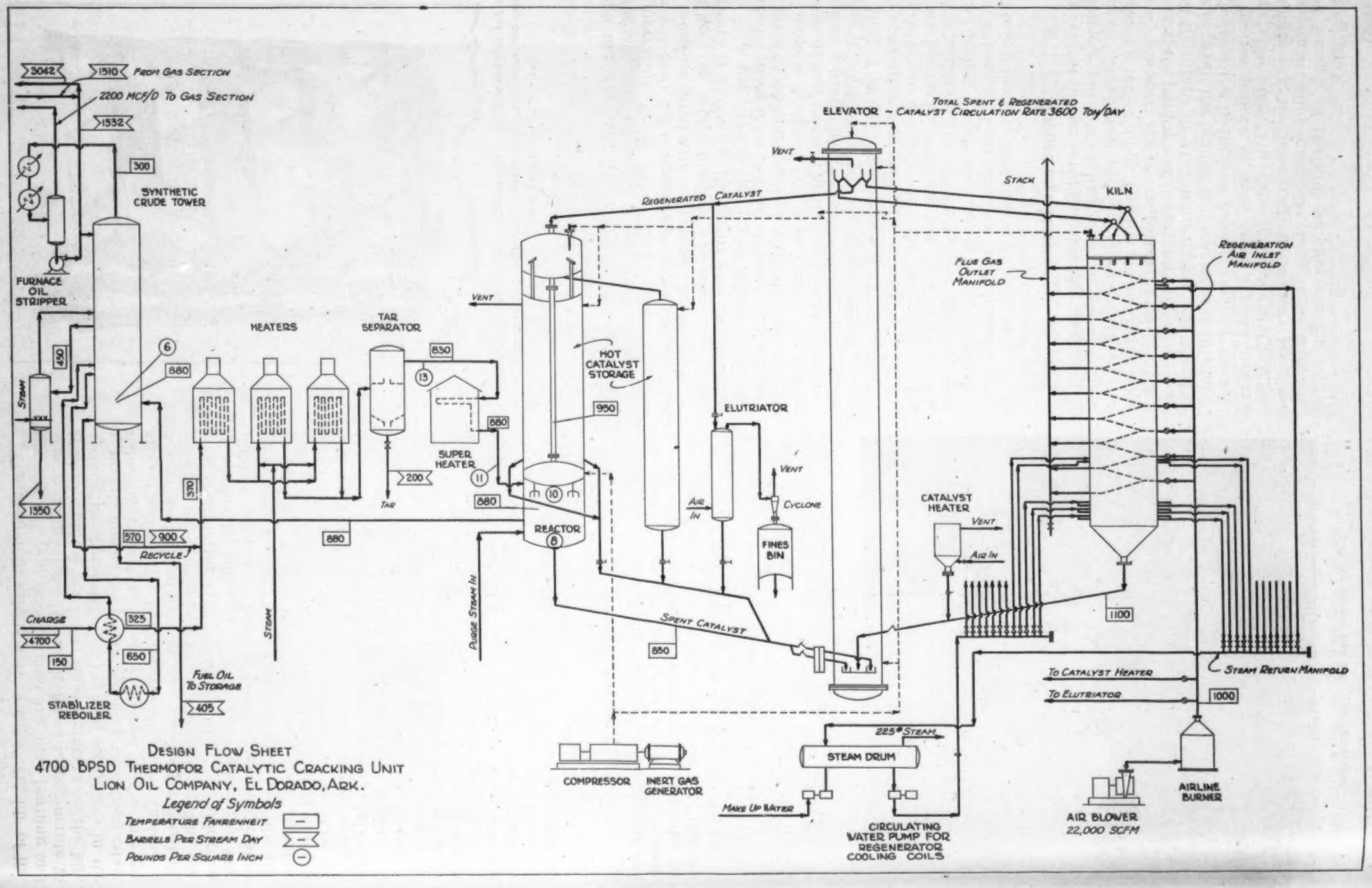
The Lion TCC unit is the second modified single-elevator Thermofor catalytic cracker to be built. Many new developments in TCC design have been included in this unit. In order to reduce new investment cost, equipment out of service which was part of former units was used as far as possible. Existing equipment utilized included all major fractionating equipment, three vaporizer furnaces, two compressors, auxiliary hot catalyst storage bin, cooling tower, control house, pump house and a portion of the pumps, exchangers and instruments.

In order to utilize an existing 12′-6″ x 64′-7″ fractionator as the new synthetic crude tower or main fractionator, it was necessary to make certain revisions. The bottom head and lower 39′-3″ of the shell was lined with ⅛-inch 11½-13 percent chrome steel, alloy type 410. The bottom eleven trays were "truss type" of the same alloy as the liner. The upper nine trays were of steel with cast iron caps and risers. The alloy trays and liner were necessary due to the high sulphur content of the charge stock. The tar stripper, a new 7′ x 25′ vessel, was completely lined and equipped with internals of 11½-13 percent chrome steel, alloy type 410. The other accumulators and drums were existing equipment of carbon steel.

The TCC process requires a two stage furnace arrangement, one stage to be a vaporizer and the second stage a superheater. Lion had available from a former unit three circular upshot gas fired iso-flow furnaces that were of sufficient size to serve as the vaporizer. Ordinarily this would be done in one heater. The entire charge first goes through one of

Heater Area Showing Two-Stage Preheaters and Superheater. Fractionation equipment in the background is synthetic crude tower on the left with tar-stripper shown between the secondary heaters.

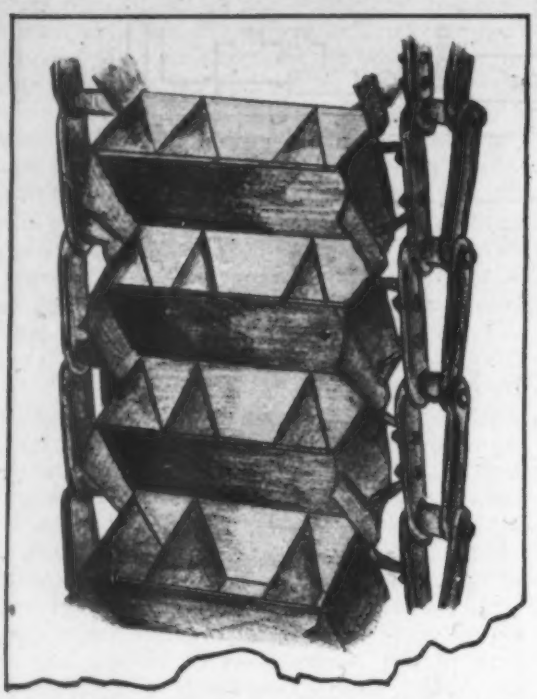
Flow Sheet for Lion Oil Company's TCC Unit



Design Flow Sheet
4700 BPSD Thermofor Catalytic Cracking Unit
Lion Oil Company, El Dorado, Ark.

Legend of Symbols

Temperature Fahrenheit —
Barrels Per Stream Day —
Pounds Per Square Inch —



Split bucket catalyst conveyor, wherein regenerated catalyst is carried in the center compartments. Spent catalyst is lifted in bucket compartments on each side.

the heaters which is equipped with carbon steel tubes and return headers. The charge is then split and goes through the other two vaporizers in parallel. These two furnaces are equipped with carbon steel tubes for temperatures below 690° F. and 2 percent chrome ½ percent moly for temperatures above 690° F. The alloy tubes are fitted with 4-6 percent chrome ½ moly cast alloy steel return headers. The total duty of the vaporizers is 34,000,000 Btu hour. The second stage heater is a gas fired super-heater which takes the overhead vapors from the tar stripper and preheats them prior to the cracking reaction. This superheater is a Lummus designed box type heater comprising two rows of roof tubes. The tube material is 2 percent chrome-½ percent moly alloy steel and the header material is 4-6 percent chrome-½ percent moly cast alloy steel. The maximum design duty of the superheater is 6 million Btu per hour.

The catalyst section, with the exception of the auxiliary hot catalyst storage bin, consists of all new material and contains all the newest developments in TCC design. These innovations in design have resulted in improved performance in addition to a reduced investment cost for the catalytic section.

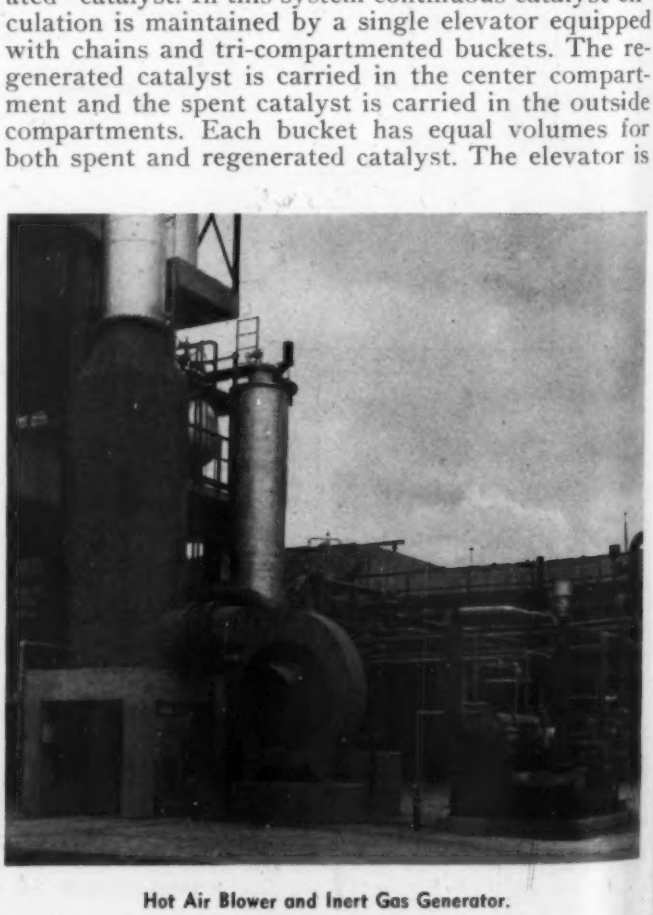
Hopper and Reactor

The reactor proper is combined with a reactor catalyst hopper and hot catalyst storage bin to form a single self-supporting cylindrical vessel 12 feet in diameter by 100′-9″ long. The top portion is the reactor catalyst hopper. The bottom part is the reactor. These two parts, the hopper and the reactor proper, are connected by means of a 50-foot catalyst seal leg. The space between the hopper and reactor and surrounding the seal leg is utilized for hot catalyst storage. The reactor proper is lined with 11½-13 percent chrome alloy steel. The vapor collector and catalyst downpipes in the disengaging section are 18-8 chrome-nickel alloy steel. The alloy steels were employed because of the high sulphur content of the charging stock.

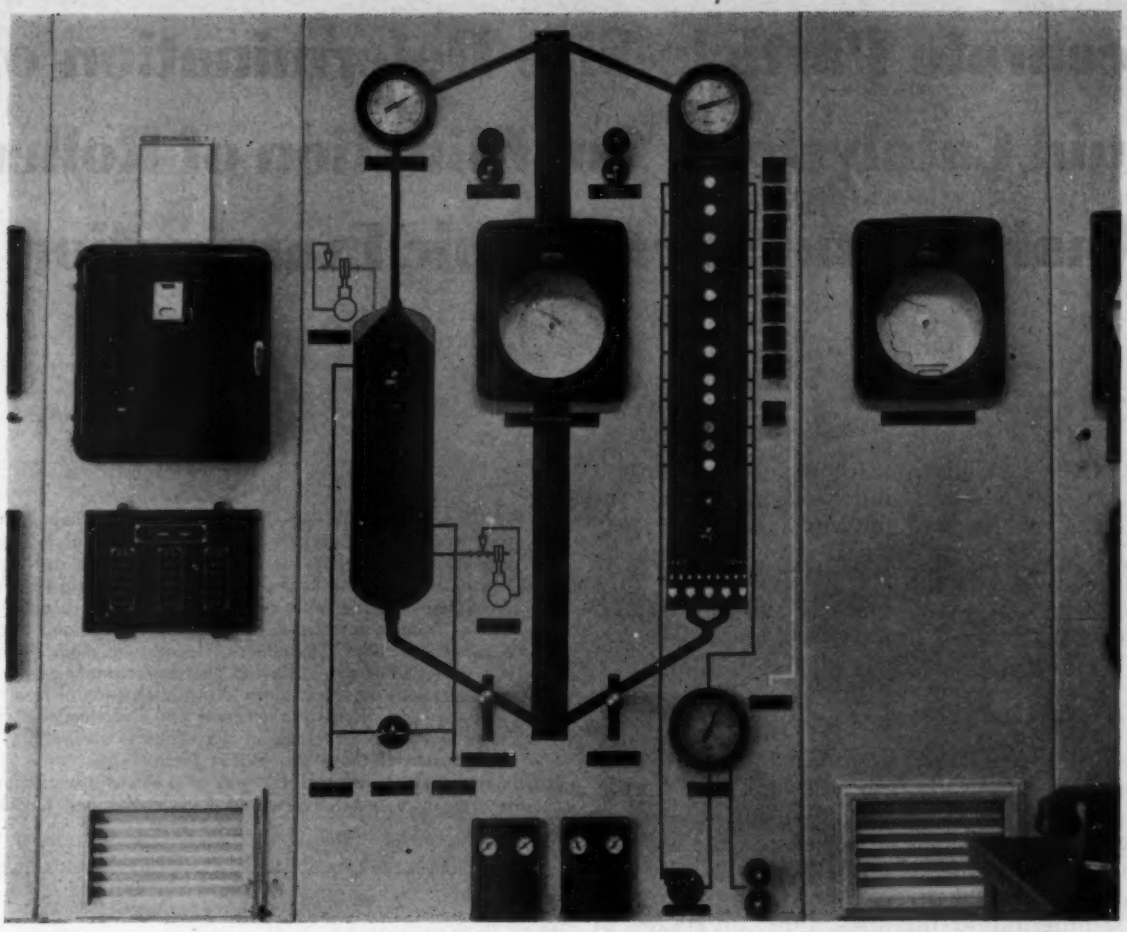
The thermofor kiln is a 10′ x 10′-6″ (inside of lining) by 95′ long vessel containing nine burning and nine cooling zones. It was designed to regenerate 75 tons of catalyst per hour. Improvements on the kiln include the elimination of cyclone separators in the main kiln flue gas stack inasmuch as it has been established that catalyst is not entrained in the flue gas stream leaving the TCC regenerating kiln. Instead of a separate kiln catalyst hopper, the hopper was built into the top of this kiln as an integral part. The vertical walls of the kiln are lined with a 6⅛″ thickness of sectionally supported tile lining consisting of 2½″ thick refractory tile backed up with high temperature insulating blocks. The roof area at the top of the kiln is lined with a four inch thickness of high temperature insulating block protected with an inner lining of 10-gauge carbon steel plate. In order to facilitate the operation of the unit the upper four air inlet dampers may be positioned from the control room panel board. The lower five air inlet dampers, which require less changing than the upper four, are manually operated. Another operating feature new to TCC units was the manifolding of all cooling coil inlets and outlets at the base of the kiln. This feature enables the operators to put a coil in or out of service at the lower level. In contrast to most TCC installations the elutriator was not attached directly to the kiln but was installed in a lower part of the structure. A small stream of hot air from the kiln inlet manifold is the source of air supply for the elutriator.

Elevator System

One of the outstanding simplifications of the unit has been the single "duplex" type elevator system. In the older TCC units two separate elevators are used, one for "spent" catalyst, and one for "regenerated" catalyst. In this system continuous catalyst circulation is maintained by a single elevator equipped with chains and tri-compartmented buckets. The regenerated catalyst is carried in the center compartment and the spent catalyst is carried in the outside compartments. Each bucket has equal volumes for both spent and regenerated catalyst. The elevator is



Hot Air Blower and Inert Gas Generator.



Part of Instrument Board Showing Catalyst Section Indicating and Control Instruments.

158 feet in length from the center of the head shaft to the center of the foot shaft. The elevator was designed for handling 150 tons of catalyst per hour, 75 tons to be spent catalyst and 75 tons to be regenerated catalyst.

A unique feature of Lion's elevator installation is the use of a steam turbine for driving the head shaft whereas all previous installations used electric motors. This installation prevents a shut down of the elevator due to power failure. On previous installations the catalyst circulation rate was determined by the calibration of a wattmeter on the motor driving the head shaft. In this case the steam flow to the turbine was calibrated against catalyst flow.

The gas section of the unit is a conventional two-stage compression plant with inter and aftercooling. The unit utilizes as the main compressor a 600 hp V-type gas engine driven compressor that handles the entire plant load. It is spared by a 400-hp horizontal gas engine driven unit that will be required only when the 600-hp unit is shut down.

The tail gas from the TCC unit gas section consisting primarily of butanes and heavier is further processed in Lion's existing vapor recovery and poly plant.

Operation of the Unit

On starting up the unit the oil section and the catalyst section were gradually heated to design temperature conditions and were initially contacted in the reactor on September 16 at a low rate of throughput. The throughput was gradually increased until design was equaled and then exceeded. The unit has been on stream only a few weeks but has already exceeded guaranteed performance and has been accepted by the management of Lion Oil Company.

Outlined below in Table 1 is a detailed analysis of one day's operation on a once-through basis.

TABLE 1

Product Yields and Inspections

October 21, 1947—Once-Through Operation

Fresh Feed Charge Stock, B.S.D. (27 API) to Reactor		4902
Conversion	Percent Vol.	59.0
Yields—On Fresh Feed:		
C_4 Free Gasoline (420° F. End Point)	Percent Vol.	49.9
Light Cycle Stock (625° F. End Point)	Percent Vol.	22.0
Heavy Cycle Stock	Percent Vol.	19.0
Butylenes	Percent Vol.	3.80
Isobutane	Percent Vol.	6.06
Normal Butane	Percent Vol.	2.22
Total C_4's	Percent Vol.	12.08
Liquid Recovery, (C_4's and Heavier)	Percent Vol.	102.98
Dry Gas	Percent Weight	5.96
Coke	Percent Weight	4.60
10 lb. RVP Gasoline	Percent Vol.	52.0
Octane Ratings of 10 lb. RVP Gasoline:		
CFR-M Clear		79.0
CFR-M +1 cc TEL		83.1
CFR-M +3 cc TEL		86.6
CFR-R Clear		90.0
CFR-R +1 cc TEL		94.1
CFR-R +3 cc TEL		97.3

		ASTM DISTILLATION, ° F.				
	°API	IBP	10%	50%	90%	E.P.
Shuler Cracking Stock	25.5	420	(1.5% at 500 ° F.)			
Topped Shuler Crude	20.8	450	(1.0% at 500 ° F.)			
Tar Separator Bottoms	17.0	...	...	...	...	...
Total Reactor Charge	27.0	410	576	696	760+	...
Motor Gasoline	64.8	90	122	208	370	420
Light Cycle Stock	24.8	454	500	540	590	624
Heavy Cycle Stock	17.3	540	646	688	760+	...

Accurate Particle-Size Determination of Fluid Catalysts by Coordination of Roller Analysis and Microscopic Examination

JOHN T. WILEY, JOE E. DELONEY, and S. WALTER DENTON
The Texas Company, Port Arthur, Texas

THIS paper was presented to a Division of Refining Symposium on Cracking Catalysts before the API meeting in Chicago last month. The study was made of the particle-size separation and distribution of fluid catalyst of sub-sieve size, using the Roller machine, with certain modifications to apparatus, and a chemical microscope with movable stage and micrometer head to check the quality of separations made. Included is a presentation of results and photomicrographs of the type of separations obtained, along with a discussion of results.

The authors discuss the accuracy and reproducibility of the Roller apparatus and some of the variables are outlined and the solution of these variables is given.

A review of different methods of particle-size distribution is given in a paper by Herbert E. Schweyer and Lincoln T. Work.[2]

AN INCREASING demand for an accurate method for sub-sieve particle-size distribution, because of appreciation of its chemical and physical importance, has made it necessary to evaluate different methods that are used to determine the particles lying with different ranges. Experience with the Roller method has brought out modifications that have tended to improve its accuracy with fluid catalyst samples.

Principles of Air Analysis

The Roller[1] method is an air-elutriation method used to separate materials into fractions of different size particles by carrying upward in an air stream those particles too small to settle against the upward velocity of the air. Elutriation methods are based on the same laws, and are affected by the same conditions, as sedimentation methods. One advantage of elutriation methods over sedimentation methods is that elutriation methods yield fractionated products which may be studied separately from the whole material, and photomicrographs can be made of each separation for permanent records if desired.

The theory or law upon which the Roller design depends is Stokes law, as follows:

$$v = \frac{10^{-8}\, g p d^2}{18n}$$

Where:

v = terminal velocity (cm per sec) in stationary fluid.
g = constant of gravitation [in cgs (metric units)].
p = density of particle, g per cu cm.
n = viscosity of fluid [in cgs (metric units)].
d = diameter of sphere in microns (1 micron = 10^{-4} cm).

Method: Roller method for particle-size distribution.

Scope: The method is intended for a rapid and accurate determination of the sub-sieve particle-size distribution of powdered materials.

Apparatus: A diagram of the apparatus used is shown in Figure 1. The mechanical parts of the apparatus are practically the same as furnished by the American Instrument Company, Silver Springs, Md.

The filter trap, with adapter shown in Figure 2, is a filter trap made in our laboratory.

The adapter is made by connecting a piece of pyrex-glass tubing with an outside diameter of about 35 millimeters and a wall thickness of about 1 mm with a standard 14-mm by 35-mm standard-taper male glass joint. The connecting tubing has an outside diameter of about 10 mm to 12 mm. The female end of the glass joint is connected to another piece

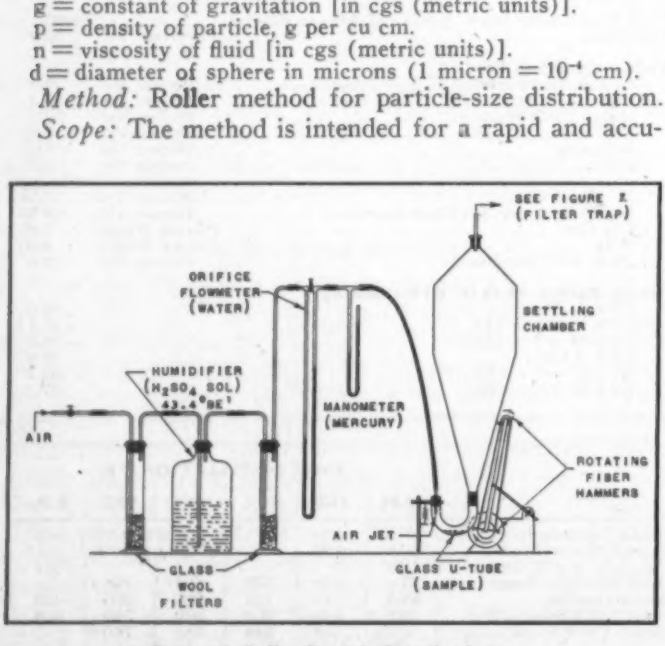


Figure 1. Roller Particle-Size Analyzer

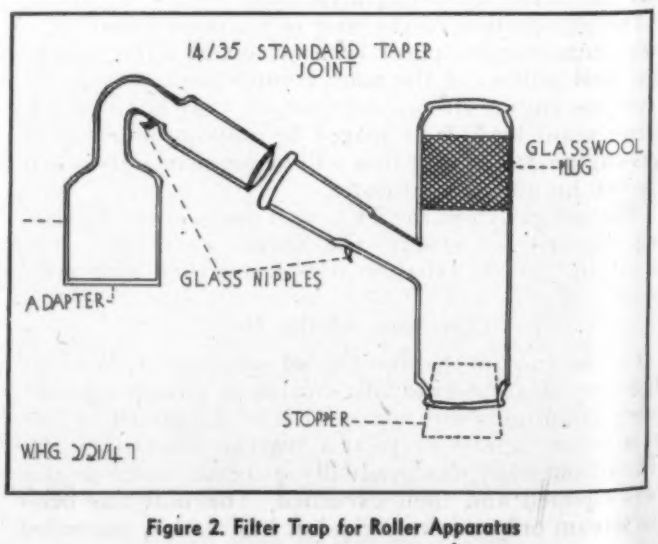


Figure 2. Filter Trap for Roller Apparatus

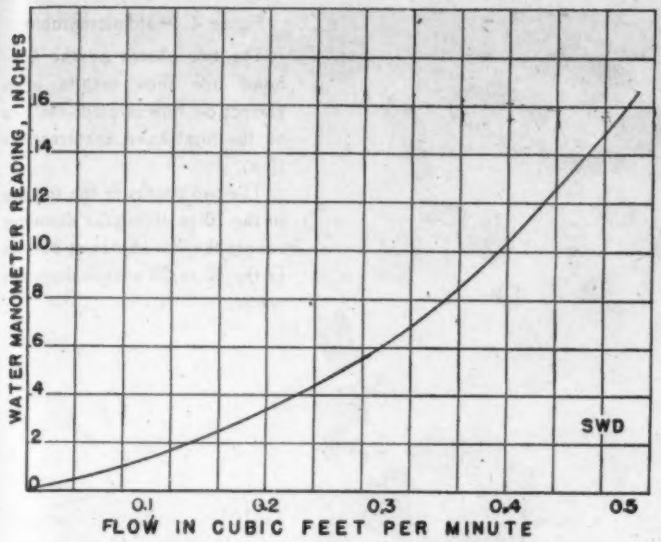


Figure 3. Air-Flow Calibration Using the 0.082-Inch Orifice

of glass tubing approximately 87 mm long, with an outside diameter of about 30 mm, as shown in Figure 2. Both ends of the glass tubing are left open, but are shaped and fine-polished to hold a rubber of cork stopper. The top end is slightly indented to hold the glass-wool plug. The adapter is connected to the separation chambers of the Roller apparatus, in the usual manner, with a piece of rubber hose. The joint between the filter trap and the adapter is held together by a rubber band or spring connected to two glass nipples on each part with a small amount of tension.

The filter trap is packed with glass wool no tighter than is necessary to catch the smallest particle of catalyst. Actually, in practice, this is not difficult. A satisfactory packed filter trap, using the 0.082 jet nozzle, produces about 1.5 inches of mercury back pressure.

Timing clock: The timing clock is an ordinary stop watch.

Club: With rubber tap on end to tap chambers.

Calibration of apparatus: The flow meter, with a suitable orifice, is calibrated for several manometer readings. A wet test meter is used for this calibration. Figure 3 illustrates typical data so obtained.

Preparation for Analysis

Samples: Unless specifically requested, samples are run in the same condition as received, i.e., samples are not redried.

The sample is well mixed, and approximately 5 grams is accurately weighed into the glass U-tube. The U-tube containing the sample, and the weighed filter trap packed with glass wool, are then attached to the apparatus, as shown in Figure 1. Then with U-tube and chamber knockers going, using the correct air flow with the right chamber, the separations

TABLE 1

Correct Air Flow with Corresponding Compacted Bulk Densities

Bulk Density (Pounds per Cubic Foot)	Air Flow (Cu. Ft. Per Min.)		Water Manometer Reading		Separation (Microns)	Diameter of Chamber (Inches)
	65	45	65	45		
.....	0.43	0.344	12.0	7.6	0-20	4½
.....	0.39	0.273	10.0	5.0	20-40	2½
.....	0.43	0.271	12.0	4.9	40-80	1½
.....	....	....	....	....	80	Residue

0 to 20, 20 to 40, 40 to 80, in that order, are collected overhead by running to an end point each time. The 80+ is the residue. The end point is determined by running a specified length of time, weighing the filter trap, and repeating the operation for successive small-time intervals until the weight-increase rate of less than 0.020 grams per minute is obtained. The approximate air-flow rate mentioned previously is given in the table. Also in the table are given the water manometer differentials produced by an orifice measured to be 0.082 inch.

Microscopic Examination

The separations obtained using a known air flow are spread on a glass slide, and are examined under the measuring microscope to determine the homogeneity and quality of the separation. Actually, in

TABLE 2

Data and Photomicrographs Showing the Relationship between Correct and Incorrect Air Flows

CORRECT AIR FLOW				INCORRECT AIR FLOW			
Micron Range	Percent	Water Differential	Cubic Feet Per Minute	Increase Air Flow (Percent)	Percent	Water Differential	Cubic Feet Per Minute
0–20	22.8	7.6	0.34	26	28.3	12.0	0.43
20–40	18.6	5.0	0.27	48	25.2	10	0.40
40–80	35.9	6.3	0.31	35	31.2	11	0.42
–80+	22.8	...	....	....	15.0	....	....

TABLE 3

Standard "A" (Used) Diakol Catalyst

Micron Range	0–20 (Percent)	20–40 (Percent)	40–80 (Percent)	80 (Percent)
1	17.7	30.7	32.4	19.2
2	16.5	30.8	35.1	17.6
3	16.2	27.6	36.8	19.4
4	15.2	30.8	35.1	18.9
5	15.1	33.1	31.2	20.3
6	15.6	33.3	33.0	18.2
7	17.0	31.8	31.2	19.9
Average	16.6	32.1	33.5	17.8

Standard No. 2 (Fresh) Diakol Catalyst

1	24.7	27.9	36.2	11.2
2	25.2	29.7	28.4	16.7
3	25.1	26.9	32.8	15.2
4	27.6	27.0	31.2	14.2

Car Composite No. 33 (Fresh) Diakol Catalyst

1	23.7	18.9	35.3	22.1
2	25.3	17.2	34.7	22.8
3	22.8	18.6	35.9	22.8
4	22.4	20.2	33.2	24.2

TABLE 4

Data Showing the Roller Method Results Vs. Sedimentation

	ROLLER			SEDIMENTATION	
CATALYST	Texas	Kellogg	Davison Chemical	UOP	Shell
Fresh					
0–20	12.0	13	14.4	10.0	24
20–40	13.3	11	13.7	13.3	13
40–80	19.2	18	25.0	28.1	30
80+	55.5	58	46.9	48.6	33
Intermediate					
0–20	14.3	16	14.8	19.6	35
20–40	18.9	21	19.2	21.9	19
40–80	21.1	21	24.3	23.8	27
80+	45.7	42	41.7	34.7	19
Equilibrium					
0–20	8.9	10	10.8	10.6	26
20–40	16.3	15	14.9	18.6	17
40–80	16.7	18	20.3	17.5	29
80+	58.1	57	54.6	46.7	28

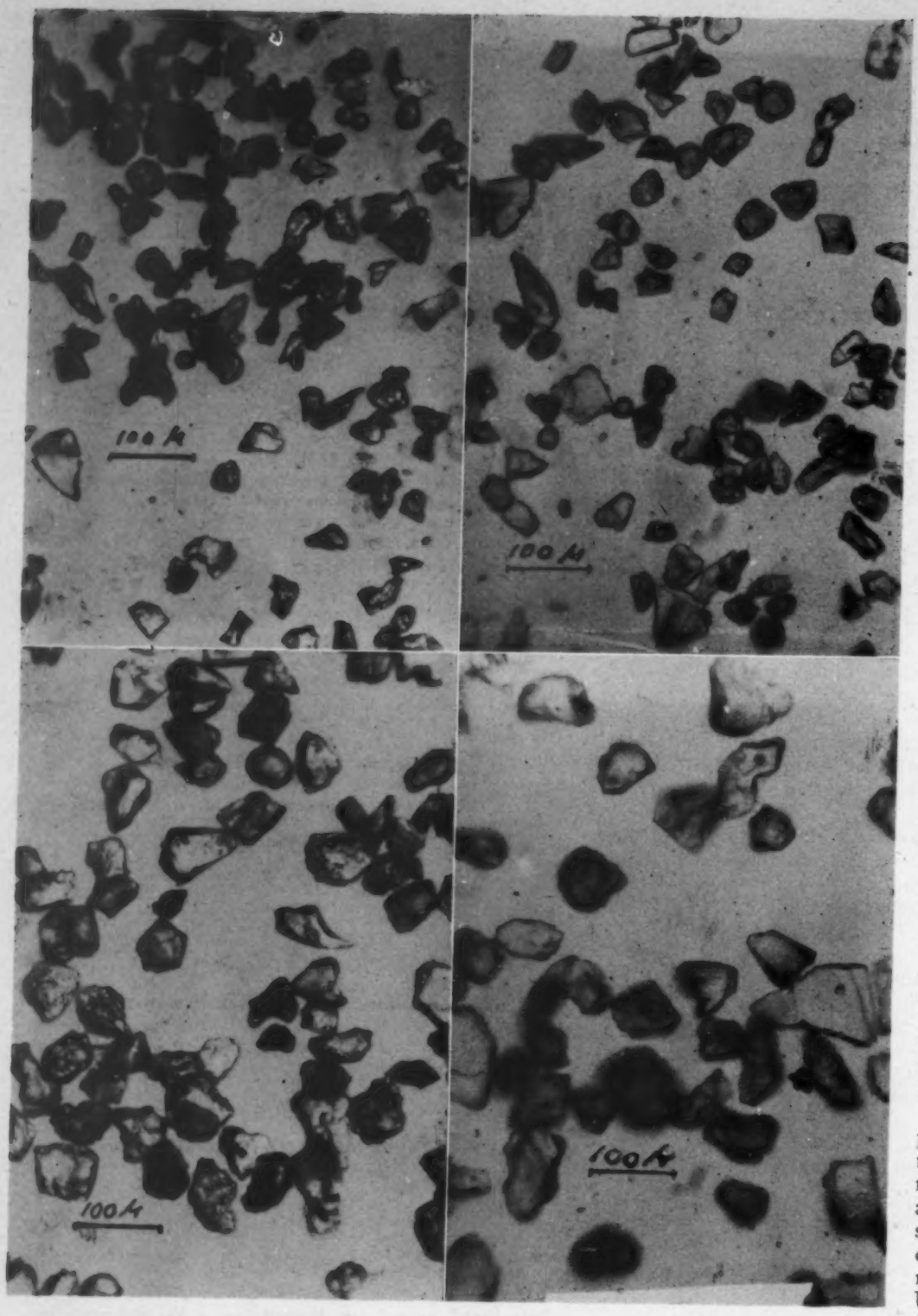


Figure 4. Photomicrographs

The two photos on the left-hand side show results when correct air flow is used; the two on the right-hand, incorrect air flow.

The two photos at the top are in the 20-to-40 micron diameter range; the two photos at bottom in the 40-to-80 micron diameter range.

practice, a separation will seldom be made without there being some overlapping. The apparatus is set up and standardized so that the error will not amount to more than 2 percent on any separation. In microscopic examination, the average diameter is used in estimating the number of particles that do not belong in a certain separation. In estimating the percentage that the particles will represent, the mass is taken into consideration—remembering that the mass varies directly as the cube of the diameter. If the analysis is not satisfactory, the air flow is increased or decreased according to the microscopic analysis until a satisfactory air flow is found.

(*Note:* If desired, more detailed information on procedure and equipment can be furnished.

The photomicrographs in Figure 5 show what we consider a good 20 to 40-micron separation on microspherical catalyst, using correct air flow along with unseparated fraction.

Table 3 indicates the reproducibility of the Roller method on two fresh Diakel catalyst samples and on one used sample. The results were obtained by several different operators over a period of several months.

The results in Table 4 were obtained by different companies when samples were sent to the companies for comparative testing.

From Table 4 it appears that the companies using the Roller method checked, whereas the ones using the sedimentation method did not — although the results obtained by Universial Oil Products Company, which used the sedimentation method, do agree to some extent with the Roller results. It is pointed out that, since the foregoing testing program, we have modified our method somewhat by using a different thimble collector (filter trap, as shown in Figure 2) and by correlating our air flow more closely with the densities—which should give even better results.

Reproducibility of the Method

The reproducibility of the method calculated from the data on the standard samples given in Table 3 is indicated to be approximately ±1.3 percent from the average. We believe this reproducibility can be obtained by careful operation and strict adherence to a

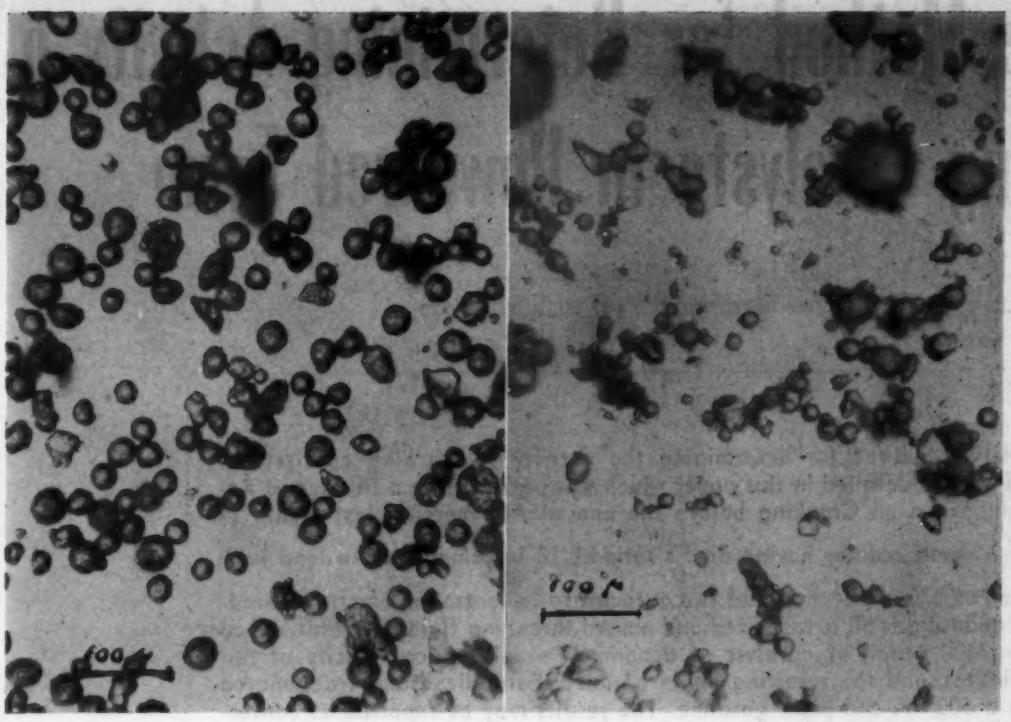


Figure 5.—Photomicrographs of Separated and Unseparated Catalyst.

The photo at left is in the 20-to-40 micron diameter range. The photo at right shows unseparated catalyst.

carefully standardized procedure, as outlined previously herein.

We believe the good reproducibility obtained is partly due to the filter trap designed in our laboratory. This glass filter trap, shown in Figure 2, eliminates the difficulty of the fast-changing weight of a paper thimble. The glass thimble is so designed that the velocity of air that carries the particle overhead is decreased when entering the chamber packed with glass wool, causing many of the particles to settle into the bottom of the tube out of the air path. This fact, and the fact that the air leaves through the top, eliminate a lot of the clogging of the thimble. Another advantage of the glass thimble is that the effective air flow can be checked, while an analysis is being run, by merely connecting a suitable means of air-flow measurement to the top opening of the thimble. In practice, this can be easily accomplished by using a rubber stopper, a piece of glass tubing, and a piece of rubber tubing.

In our work, no correction of the air flow was found necessary for the small variations in barometric pressure, temperature, and mercury back pressure. Also, inasmuch as the types and percentages of the samples were similar, the compacted bulk density was satisfactory for air-flow correlation. In extreme cases it may be necessary to use the less easily determined absolute density.

Discussion of Results

The foregoing data show that accurate reproducible results can be obtained by the Roller method if the precautions listed hereinafter are observed:

1) Correlate density and air flow.
2) Check air-flow calibration frequently.
3) Check the Roller equipment for leaks frequently.
4) Use the glass filter trap with adapter, as shown in Figure 2.
5) Check the end point on each separation.
6) Examine each separation under the microscope for quality of separation.

Factors affecting the results, which we have encountered only in extreme cases, and which cannot be controlled effectively in the analysis, are as follows:

1) If there is a large percentage concentration of particles around the end point, the end point would become "supercritical"; i.e., small variations in the air flow would affect the separation or fraction percentage appreciably. However, when examined under the microscope, it may be very difficult to detect the change. Good separation depends upon uniform, or at least, representative, distribution around the end point.

2) Another factor that might affect the results would be if the density of the catalyst being analyzed were not uniform or almost so; then it would be difficult to obtain a good separation.

3) If the shape of the catalyst is not spherical, the mean diameter is difficult to determine under the microscope. As the air-rate adjustment is dependent upon this examination, care should be used in making the examination.

Conclusions

The Roller method with carefully standardized apparatus, using the modified filter trap and close correlation of air flow with density, is a convenient rapid method that gives reproducible results when used along with a measuring microscope. Photomicrographs can also be taken on any separation for a permanent record, if desired.

An experienced operator can make four Roller analyses, and check the results under the microscope, within eight hours.

Acknowledgment

The writers wish to acknowledge the helpful suggestions made by various members of Standard Oil Company of Louisiana at Baton Rouge in the development of the experimental method; the photomicrographs by H. D. LeRosen of our laboratory; the results and experience accumulated through the efforts of various members of The Texas Company PAW Analytical Laboratory; and the helpful suggestions of W. S. Palmer in the writing of the paper.

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Bench-Scale Method for Determining Activity of Cracking Catalysts in Powdered Form

HUBERT McREYNOLDS
The Texas Company, Port Arthur, Texas

A QUICK method for determining the activity of cracking catalysts in powdered form is detailed in this paper which was presented to a Division of Refining Symposium on Cracking before the annual API meeting last month in Chicago. A small reactor having an $\frac{L}{D}$ ratio of 18 is immersed in a lead bath maintained at 920° F. by means of two automatically controlled electrical windings. A standard gas oil is vaporized and passed through a fluidized catalyst bed containing 400 grams of catalyst at a nominal weight space velocity of two for 30 minutes, and the resulting liquid product is stabilized and fractionated to obtain a 400°-F.-end-point gasoline. The results may be reported as conversion, i.e., 100 minus percent by volume of gasoil, but D + L† (percent distilled plus loss) is preferred, as only the distillation of the total cracked product is involved in its calculation. The Texas D + L has been correlated with the widely used Jersey D + L, and this correlation is included.

This method also provides data from which the gas and carbon-producing tendencies of catalysts may be determined. These factors are of interest, because they generally indicate the degree of contamination.

The reproductibility of the test is good, as indicated by data from 29 runs over a standard catalyst. The data indicate, for example, that for 2 determinations the true value is expected to fall within the following limits, 95 times out of 100:

	Percent
Conversion (100 minus percent by volume of gasoil)	1.84
Texas D + L	1.20
Percent by weight of carbon (basis, fresh feed)	0.123
Gas production, cubic feet per barrel of feed	43.4

† D + L = distillate plus loss.

CRACKING catalysts used in commercial catalytic-cracking operations lose activity with continued use. Obviously, it is very desirable to know the activity of the catalysts in service at any stage of deactivation, inasmuch as activity is one of the variables governing the conversion of feed stock to useful products. From an operational standpoint, a measurement of catalyst activity at frequent intervals permits the operator to follow the activity decline of a catalyst and adjust operating conditions to maintain conversion at the desired level. This may be done on a day-to-day basis by means of surface measurement by potassium-hydroxide absorption or any similar method, but there must be periodic determinations of the true activity of the catalyst in order for the relative surface measurements to have any real meaning. Changes in operating conditions or feed stock, for example, may result in a difference in product distribution which would change the catalyst activity, i.e., potassium-hydroxide index relationship, making the determination of the actual catalyst activity necessary.

Many different methods of measuring and expressing the activity of cracking catalysts are currently in use. It is generally accepted, however, that a catalyst-activity test method should indicate the gasoline-producing ability of a catalyst as well as its tendency to form carbon and gaseous products. These latter factors are important, because they indicate the extent to which the catalyst may have been contaminated in use.

It is the consensus that an ideal fluid catalyst-activity evaluation method should combine, as its chief characteristics, the following points: It should be capable of testing fluid catalysts without the necessity for pelleting. Operating temperatures and temperature gradients should be capable of being measured accurately. The process periods should be long enough for fluctuations about a steady state to average out. Sufficient charge stock should pass through the unit to insure ample product for testing. The conditions under which the test unit separates products into liquid and gas, such as receiver and knock-back condenser temperatures, must be definite and reproducible. Distillation of the liquid product should be done by use of an automatic Hempel column to give a precise measure of conversion. A definite pretreatment or conditioning of the catalyst, to

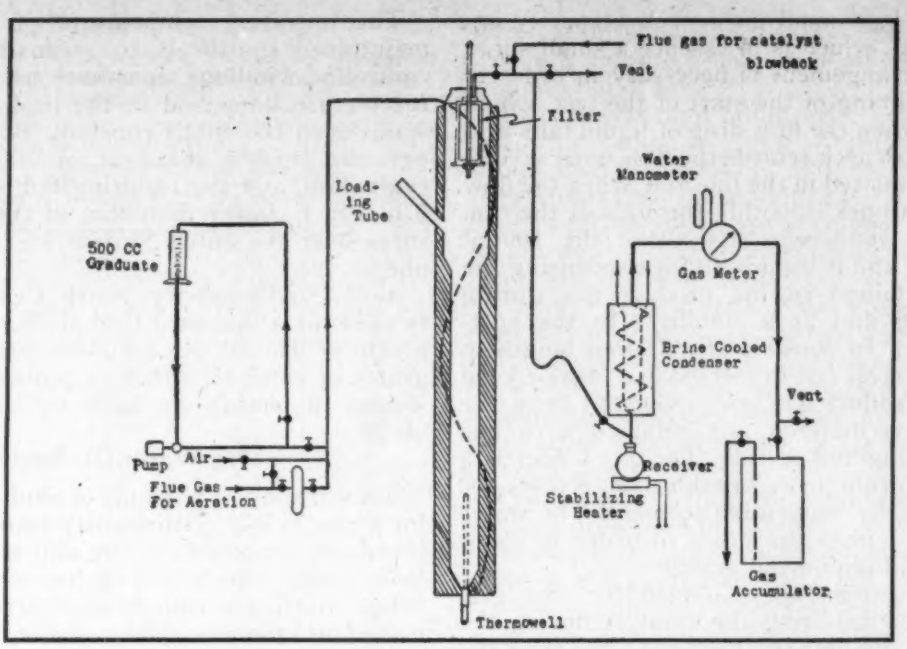

Figure 1. Schematic Flow Diagram

put it into a "standard state" before testing, should be adhered to.

The catalyst test method in use at Port Arthur satisfies these requirements to a large extent. Briefly, the method consists of charging a standard gas oil through 400 grams of powdered catalyst at about 920° F. for 30 minutes, the resulting liquid being stabilized and fractionated to yield 300°F.-end-point gasoline. The activity of the catalyst may be reported in terms of conversion (100 minus percent by volume gasoil) or the percent distilled plus loss (D + L) on the total liquid product. This method may be correlated with other evaluation methods, for example, the Jersey method, to express catalyst activity in other terms.

Description of Equipment

A schematic flow diagram of the test unit is shown in Figure 1. It will be noted that it consists of a charge system, an upflow fluidized fixed-bed reactor, a brine-cooled condenser, a gas meter, and a gas-accumulating drum. When this equipment is used for activity determinations, a calcined catalyst is loaded into the unit, and a few minutes are allowed for the catalyst to come to temperature. The run is then made without any further delay. As soon as the run has been completed, the catalyst is purged with flue gas and drawn from the unit. No provision is made to regenerate or reuse the catalyst; if additional runs are required, they are made on fresh portions of the same catalyst. The carbon content of the spent catalyst may be determined directly by the Leco, or any similar, method.

The reactor is made from a 36-inch section of stainless-steel tubing. The larger 18-inch disengaging section above the reaction is made from 3-inch tubing of the same material. The cone at the bottom of the reactor is carefully shaped to give an angle of 30° from the perpendicular to the wall in order to avoid catalyst holdup when the spent catalyst is withdrawn from the bottom of the reactor. The bottom thermowell may be removed for this purpose.

A small amount of flue gas is routed through the reactor before catalyst is loaded, in order to keep the preheater inlet clear. Catalyst is loaded through the side connection in the disengaging space. This section of 1-inch tubing is filled with a piece of bar stock during operations to keep catalyst and vapors out of this line. After the catalyst has been loaded, the flue-gas rate is increased to maintain the catalyst in a fluidized state until the charge has been started through the heater.

Catalyst is removed from the vapors, leaving the disengaging section by means of an Aloxite filter tube 6 inches long and 1⅜ inches in outside diameter. This filter has a permeability of five, and can be used for several months before plugging is experienced. It is held in place by an Invar rod drilled to receive a thermocouple which may be used to measure the vapor-outlet temperature. Invar was selected because of its low coefficient of expansion, making possible a tighter fit at the top and bottom asbestos gaskets which line the steel heads holding the filter tube.

The entire reactor and disengaging assembly is immersed in a 6-inch-diameter lead bath which is heated by two electrical resistance coils. The current is automatically controlled to maintain lead-bath temperatures at 920° F. at points near the middle of the reactor and at about the center of the disengaging section. The charge heater, which consists of a 6-foot section of ⅜-inch stainless-steel tubing, is wound around the reactor assembly. It enters the lead bath at the top, and extends to the preheater inlet in the cone at the bottom of the reactor. This heater is long enough to assure a constant heater outlet temperature of 920° F. at the charge rates normally maintained, but is small enough to avoid excessive thermal cracking.

The standard gasoil feed stock is pumped from a 500-milliliter graduate into the charge heater by means of a small horizontal proportioning pump. These pumps have been found to be dependable, particularly when equipped with double checks, and are easily adjustable to regulate the pumping rate.

Accurate Timing of Start Obtained

Oil vapors leave the reactor through a filter, and enter a 500-ml. product receiver flask by means of a ⅛-inch transfer line. This line serves as an air-cooled condenser and contains, immediately upstream from

the receiving flask, a small gooseneck (capacity approximately 3 ml), which is in essence a small surge chamber. This arrangement is necessary in order to obtain accurate timing of the start of the test, which starts officially when the first drop of liquid falls into the flask. The gooseneck retards the flow until several drops have accumulated in the line and, when the flow does start, it continues smoothly throughout the run.

A knock-back condenser sits above the 500-ml product receiver, and is used both for condensing the light liquid contained in the product gas during actual processing and as a stabilizer in the post-processing period. In consists of a 16-foot length of ⅜-inch stainless-steel coil immersed in a brine-cooled kerosine bath. Product gas goes overhead from the condenser, and is metered and retained in a gas holder for sampling and testing. The gas holder is a 30-gallon metal drum with suitable connections for retaining the gas by water displacement. The water may be syphoned from the drum in order to minimize pressure buildup on the system.

Suitable connections are provided for flue gas, which is used for fluidizing the catalyst during the warm-up period and blowing back the filter after the run has been finished. Air connections are available for periodically decoking the heater.

Fractionating equipment consists of a modified Hempel column packed with 3/16-inch stainless-steel Fenske packing. The packed column is connected with the 1-liter pyrex still pot by means of a ground-glass joint. Heat is supplied to the still pot by a 750-watt heater controlled by a variable transformer. Cooling is supplied by a brine-cooled condenser coil, and is maintained at about 32° F. No cooling is supplied to the receiver graduate, as the liquid has been well stabilized and losses of light ends are negligible. Weight recoveries on the distillation are normally in the range from 98 to 100 percent.

Testing Procedure

Used catalyst is pretreated by calcination at 1000° F. in a stream of dry air for two hours for carbon removal. Prior to charging to the unit, the catalyst is heated in an oven to approximately 250° F. for ease of loading. The same procedure is employed with fresh catalyst, but calcination temperature is limited to 850° F.

It has been found that catalysts containing more than 20 percent 0 to 10-micron particles or catalysts having a mean particle size lower than 30 microns tend to agglomerate, resulting in channeling of the vapor stream through the catalyst bed. This condition is circumvented by testing samples from which a portion of the fines has been removed by air elutriation. Investigation of a possible difference between the activity of the coarse and fine portions has been confined to a study of the potassium-hydroxide indices obtained on these materials, as no satisfactory method has been developed for directly evaluating the activity of the fines. A statistical study of the data has shown that a significant difference exists in the potassium-hydroxide index of the coarse and the fine portions of the same catalyst, although there is no consistent difference; i.e., values of fines may be higher or lower than on the heavier fraction. This difference is generally quite small, being in the order of 0.02 potassium-hydroxide index units, corresponding to approximately 0.5 Texas D+L units. It is considered that a difference of this magnitude, although statistically significant, is of no practical importance.

The operating temperature of the lead bath is maintained at 920° F. by means of the electrically controlled windings. Inasmuch as the preheater and reactor are immersed in the lead, temperatures are maintained assentially constant. The reactor-bed temperature reflects the heat of cracking during the evaluation, and the resulting temperature is usually 5 to 10° F. lower than that of the lead bath. Pressures over the entire system are essentially atmospheric.

A 35.7°-API-gravity North Central Texas gasoil is used for a standard feed stock, and is charged at a rate of 925 ml per hour for 30 minutes over 400 grams of catalyst, which is equivalent to a nominal weight of charge per hour per weight of catalyst of 2.0.

No Warm-Up Required

No warm-up of the unit is required in preparation for a run, as it is continuously maintained at 920° F.-lead-bath temperature. The unit is merely tested for leaks under 1 inch to 2 inches of mercury pressure before starting a run. A small stream of flue gas is routed into the unit to keep the bottom tee and heater clear, and the catalyst bed is fluidized while the catalyst is being loaded. This rate is increased to 5 to 6 cubic feet per hour after the catalyst is in place, and is maintained until fresh feed charge has been established.

The start of the test period is taken as the time that the first steady stream of oil flows into the receiver. At this point the line which bypasses the gas-collector drum is closed and the valve into this drum opened so that the overhead gas will be collected. The following readings are obtained at 5-minute intervals during the test period: top lead, bottom lead, reactor catalyst bed, knock-back condenser, and pressure (inches of water) on the gas meter. The disengaging-space outlet temperature and gas-meter readings are read at the beginning and end of the test period. The charge rate is checked at 10-minute intervals.

Exactly 30 minutes after the start of the test period, the receiver flask and gas-recovery system are blocked off from the reactor. Flue gas is routed to the reactor at a rate of 5 to 6 standard cubic feet per hour to strip oil vapors from the catalyst. After 15 minutes of stripping, the catalyst is withdrawn from the reactor, cooled, and weighed. The filter and disengaging space are blown free of catalyst with flue gas, and the preheater is cleaned of any residual oil by passing flue gas through it during the purge period. About once a week, the preheater is burned free of any coke with air.

The knock-back condenser is held below 32° F., preferably not less than 25° F., during the test period. About half-way through the run, the heating element is put in position to apply heat to the bottom of the product flask. The temperature of the condenser is allowed to rise to 32° F., and boiling of the liquid product is continued until no more gas is evolved at these conditi s. This technique of stabilization results in essential debutanization of the total liquid, with part of the pentanes going overhead. A typical example of the hydrocarbon split is shown as follows:

COMPONENT	Fresh Feed (Percent by Weight)	
	In Gas	In Liquid
Butanes	3.9	0.1
Pentanes	1.4	1.7

TABLE 1

The Texas Company D + L, Conversion, Percent by Weight of Carbon, and Gas Make Obtained on Standard Low-Activity Catalyst

Test No.	Texas Company D+L (Percent by Volume of Gasoline)	Conversion (100 Minus Percent by Volume of Gasoil)	Carbon (Basis, Fresh Feed) (Percent by Weight)	Gas Make (Cubic Feet Per Barrel)
1	17.2	21.9	0.61	168
2	17.3	21.2	0.64	161
3	18.9	24.6	0.62	204
4	18.4	23.9	0.58	230
5	18.7	23.5	0.65	207
6	18.2	23.0	0.60	206
7	18.3	22.3	0.65	202
8	17.7	23.7	0.63	182
9	19.6	22.8	0.57	201
10	19.5	25.0	0.56	247
11	18.3	23.5	0.67	211
12	17.4	21.9	0.72	226
13	17.6	21.6	0.59	195
14	19.7	24.0	0.63	234
15	16.7	21.1	0.67	201
16	18.8	24.4	0.63	241
17	17.8	24.8	0.82	259
18	16.9	23.4	0.80	236
19	17.3	22.9	0.77	244
20	17.8	22.4	0.70	223
21	18.3	25.8	0.68	272
22	17.4	23.7	0.78	293
23	17.3	22.0	0.70	217
24	18.3	23.8	0.56	220
25	19.2	25.9	0.90	259
26	16.8	23.9	0.62	213
27	17.6	22.8	0.72	250
28	17.6	24.6	0.78	250
29	18.0	23.7	0.63	240
Average	18.0	23.4	0.67	224
Number of runs	29	29	29	29
Standard deviation	0.83	1.27	0.085	30

After the stabilized liquid product has cooled, it is weighed and the API gravity is obtained. It is then fractionated on a modified Hempel apparatus to a 390°-F. vapor temperature, which results approximately in a 400°-F.-end-point gasoline by ASTM distillation. The boil-up rate is controlled to give about 3 ml per minute of gasoline overhead.

API gravities and ASTM distillations are obtained on the gasoline and gasoil fractions. Specific gravity (Acme) and mass-spectrograph analysis for light hydrocarbons are obtained on the overhead gas. Where a more complete hydrocarbon breakdown is desirable, the total liquid is also analyzed for the light hydrocarbons through pentanes. Catalysts are tested in duplicate for percent carbon.

Reliability of Results

The data thus obtained are used to estimate the catalyst activity, which may be expressed in two ways:

1) *Percent conversion,* which is 100 minus percent by volume of gasoil yield; basis, fresh feed; i.e.: gasoline, gas, and carbon plus loss. Fresh Diakel catalyst results in a conversion of about 55 percent on this basis.
2) The Texas Company D + L, which is 100 minus percent by volume of gasoil yield; basis, total liquid product; i.e.: 400°-F.-end-point gasoline produced from the total liquid plus distillation loss. Fresh Diakel catalyst has an activity of approximately 40 percent on this basis.

The reproducibility of catalyst evaluation data, on the basis of the Texas D + L and conversion, are shown in Table 1. Also shown are the percent by weight of carbon on the basis of fresh feed, and the gas expressed as cubic feet per barrel of fresh feed. These data were obtained from 29 runs made at intervals over a period of seven months on a standard low-activity catalyst typical of plant equilibrium catalysts from commercial operations. On the basis of the estimated population standard deviations of a single determination shown in Table 1, the limits are given within which the true average may be expected to fall 95 times out of 100.[1]

Two observations are made on each catalyst tested, and are expected to check within 2.5 percent of The Texas Company D + L. If they do not fall within these limits, additional values are obtained—in which case all observations are averaged. Weight recovery on any evaluation run must be within the limits of 97 to 102 percent to be used.

Carbon determinations are based on the Leco method for analysis of percent carbon. The overall reliability of the method, and the system of sampling, have been tested; and the standard deviation basis of 82 samples averaging 1.36 percent carbon was found to be 0.069 percent, which is 5.1 percent of the observed value.

Typical results from the modified Hempel distillation are presented in Table 3.

Although considerable variation is experienced in naphtha end point (basis, ASTM distillation) naphtha yield is affected relatively little.

Correlation With Other Methods

Either or both of the two methods of expressing catalyst activity can be correlated with other methods. For example, the more commonly used Jersey D + L[2] activity has been correlated with the Texas D + L, and is shown in Figure 2. This correlation is based on 17 catalysts tested by each method; eleven of these catalysts were tested in cooperation with the special panel of testing of the catalytic-cracking subcommittee of PAW Aviation Gasoline Advisory Committee.

Gas- and Carbon-Producing Factors

These factors are defined as the ratio between gas and carbon production of an unknown catalyst relative to the yields obtained on an uncontaminated catalyst of the same activity. A base correlation was

TABLE 2

Number of Observation "N"	Values of "$a\sigma_x$" in $X \pm a\sigma_x$ for 95 Chances in 100			
	Texas Company D+L (400 °F. End-Point Gasoline Plus Loss)	Conversion (Percent by Volume of Gasoil)	Carbon (Basis, Fresh Feed (Percent by Weight)	Gas (Cubic Feet Per Barrel)
1	1.70	2.60	0.174	61.4
2	1.20	1.84	0.123	43.4
3	0.98	1.50	0.101	35.5
4	0.85	1.30	0.087	30.7
5	0.76	1.16	0.078	27.5
6	0.69	1.06	0.071	25.1
	$\frac{(2.048)\ (0.83)}{\sqrt{n}}$	$\frac{(2.048)\ (1.27)}{\sqrt{n}}$	$\frac{(2.048)\ (0.085)}{\sqrt{n}}$	$\frac{(2.048)\ (30)}{\sqrt{n}}$

TABLE 3

Typical Results from Modified Hempel Distillation—Cut Point, 390° F.

	Yield of Gasoline (Percent by Volume)	ASTM End Point of Gasoline	ASTM Initial Boiling Point of Gasoil
	19.1	406	460
	19.0	392	460
	18.8	392	466
	18.5	408	466
	19.1	390	480
	19.3	394	472
	19.1	404	464
	18.8	406	466
	19.1	408	464
Average	19.0	400	466
Number of points	9	9	9
Standard deviation	0.18	7.8	6.3
"$a\sigma_x$" in $X \pm a\sigma_x$ for 95 chances in 100 basis; 2 observations	0.29	12.8	9.9

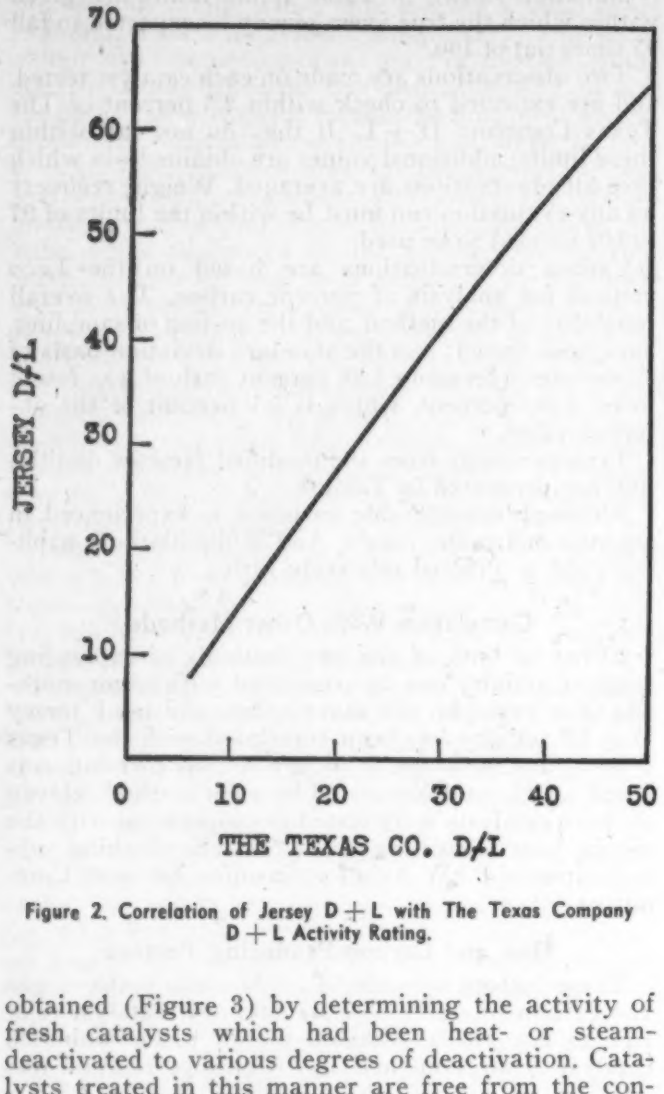

Figure 2. Correlation of Jersey D + L with The Texas Company D + L Activity Rating.

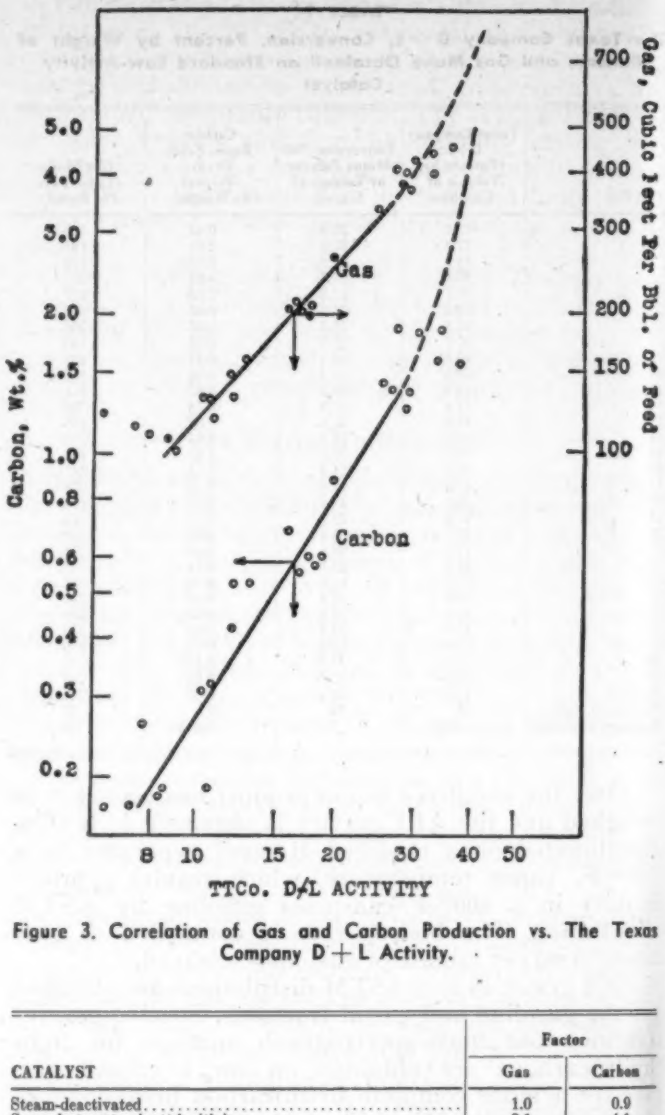

Figure 3. Correlation of Gas and Carbon Production vs. The Texas Company D + L Activity.

obtained (Figure 3) by determining the activity of fresh catalysts which had been heat- or steam-deactivated to various degrees of deactivation. Catalysts treated in this manner are free from the contaminants received in actual service such as iron, nickel, etc. Basically, a fresh catalyst will have a factor of 1.0, and catalysts having ratios greater than 1.0 are considered to be contaminated. Figure 3 was developed from data obtained on 26 runs employing 13 3A Diakel catalysts ranging from fresh to almost completely deactivated.

The effect of contamination on the gas- and carbon-producing factors of several catalysts may be illustrated by data in the accompanying table.

The steam-deactivated catalyst was not contaminated, whereas the second of the foregoing catalysts was a sample of the same material which had been impregnated in the laboratory with the equivalent of 0.3 percent by weight of ferric oxide. Commercial equilibrium catalysts (A) and (B) are examples of typical contaminated and uncontaminated catalysts.

CATALYST	Factor	
	Gas	Carbon
Steam-deactivated	1.0	0.9
Same, but contaminated with iron	2.1	4.1
Commercial catalyst (A)	1.1	1.1
Commerical catalyst (B)	1.9	3.2

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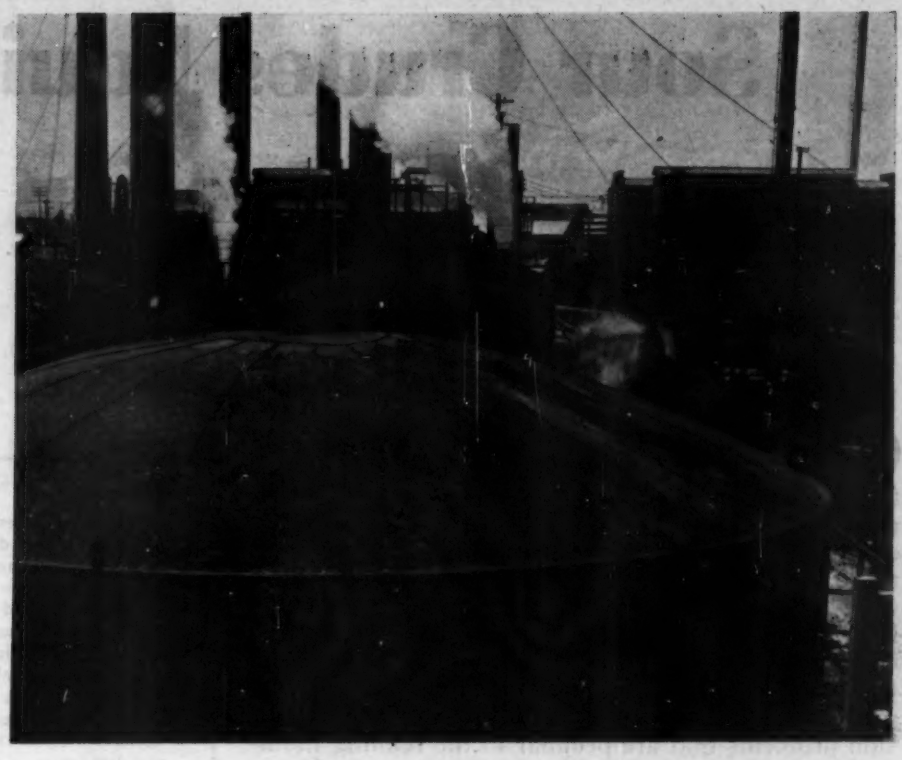
[1] *ASTM Manual on Presentation of Data*, 3rd printing, American Society for Testing Materials, Aug. (1940).

[2] M. E. Conn and G. C. Connolly, *Ind. Eng. Chem.* 39, 1138-43 (1947).

Aluminum Used In Combatting Tank Corrosion

GORDON W. ALEXANDER
Associate Editor



Typical Corrosion Failure of Roof Plates on Rundown Tanks Containing West Texas Distillate

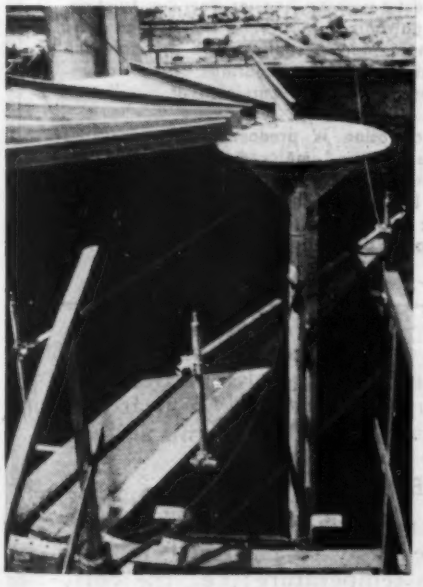
EVER present around the refinery is the enigma of corrosion. Regardless of the tremendous amount of information on corrosion, particularly in regard to the corrosion of most of the common materials, it is a stubborn fact that the corrosion resistance of any material for a particular application can be definitely determined only by exposing that material to the existing conditions.

Eastern States Petroleum Corporation, Houston, is combatting the problem of the corrosive reaction of hydrogen sulphide on tank roofs and their supporting structure by use of sheet and tubular aluminum.

The tanks in question are rundown tanks for gasoline cut from the thermal cracking unit. The charge to this unit is a typical sour West Texas crude containing a comparatively large amount of sulfur. When processed, the crude produces a corresponding amount of hydrogen sulphide which reacts with the metal to form iron sulphide. This reaction, occurring from the inside of the tank, is most difficult to combat, and it was only a matter of time until tank roofs and their supporting had to be replaced.

Because aluminum is resistant to hydrogen sulphide and free sulfur it was chosen for use in replacing the tank roofs. Thus all metal exposed to the inside of the tank with the exception of the tank shell is aluminum. The rivets are aluminum and are used throughout on the roof fabrication. The center column support is welded. Lead is used as a gasket to make the metal joints vapor tight.

The center column consists of two 4½-inch o.d. 2S 1/2 hard aluminum pipes as the supporting element. An aluminum plate ½-inch thick provides the bearing for the center column on the tank bottom. At the top of the center column a circular plate serves as a hub to which the roof sheeting is riveted. The roof is fabricated in the shop in 90 degree segments and then fitted to the tank. Four-inch channel steel is used for rafters. The rafters are placed on the top side of the aluminum roof sheeting and are spaced at 18-degree intervals. The aluminum sheeting is riveted to the rafters with aluminum rivets.



Crown Detail of Center Column

Base Detail of Center Column

A Completed Roof

Sour Crudes Equipment Protection

E. Q. CAMP,
Humble Oil & Refining Company, Baytown, Texas

THIS paper, presented at the API Division of Refining session in Chicago, is a discussion of some of the major corrosion problems experienced in refining sour crudes. Corrosion at temperatures below about 500° F. is classified as low-temperature corrosion, and that at temperatures above about 500° F. as high-temperature corrosion. The more important factors involved in corrosion in these temperature ranges are turbulence, pressure, and the presence of corrosive concentrations of such materials as reactive sulfur compounds, inorganic salts, naphthenic acids, oxygen, and water.

The corrosion that occurs in the storage, transportation, and distillation of sour crudes and sour distillates is predominately low-temperature corrosion. The life of tanks in sour-crude service is increased by gunite coatings, coatings of paints and plastics, caustic washing, heavier steel construction, use of aluminum and galvanized iron roofs, and wetting roofs with crude oil. Of these, gunite coatings are probably the most satisfactory. Neutralizers, gunite coatings, and corrosion-resistant alloys are used to control corrosion of distillation equipment.

High-temperature corrosion is predominant in cracking service. The protective measures employed in this type of corrosion include extensive use of corrosion-resistant alloys, gunite coatings, and linings of alloys, as well as the use of neutralizers in charge and product.

The role of sulfur as an inhibitor of two types of corrosion is discussed.

CORROSION is one of the major problems confronting the designer and operator of refinery equipment. It has been estimated that the losses due to corrosion in the refining industry alone are equivalent to approximately 1 cent per gallon of gasoline produced.[1] These enormous losses occur as a result of the large number of corrosion problems that are experienced in the refining industry. These corrosion problems include essentially all corrosion common to other industries and, in addition, a number of corrosion problems that are peculiar to the refining industry. The corrosion problems peculiar to the refining industry are primarily those which arise as a result of the presence of corrosive sulfur compounds and of chlorides of magnesium and calcium in the crude oils. The major offenders in this respect are the high sulfur crudes, frequently referred to as sour crudes, which include West Texas and Panhandle. This paper presents a discussion of the major corrosion problems experienced in processing sour crudes at the Baytown refinery of Humble Oil & Refining Company, and of the methods employed in mitigating these corrosion problems at this refinery. This paper is limited to a discussion of the major corrosion problems experienced in processing sour crudes, because the minor problems encountered are so many that it would not be feasible to discuss them in one paper.

The corrosion experienced in processing sour crudes has been discussed in detail in a number of papers presented on this subject during the past several years.[2 to 22] Although the present paper is limited primarily to the experiences with corrosion in one refinery, and to corrosion mitigation in processing sour crudes, it is emphasized that these experiences closely parallel those in other refineries in which sour crudes are processed; and, as a result, many of the problems discussed and the solutions presented in this paper have already appeared in the literature that has been published on this subject.

Classification of Corrosion

The corrosion of refinery equipment in which sour crudes are processed may be classified, for convenience, according to temperature; i.e., the corrosion which occurs at low and high temperature.[2] Corrosion

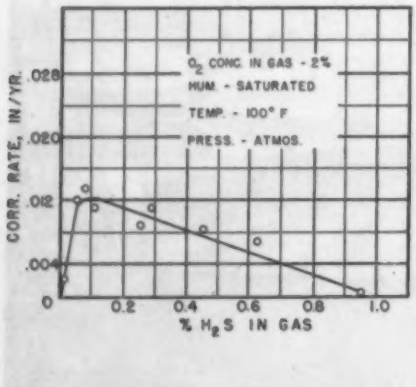


Figure 1. Effect of H_2S Concentration on Gaseous Corrosion of Steel.

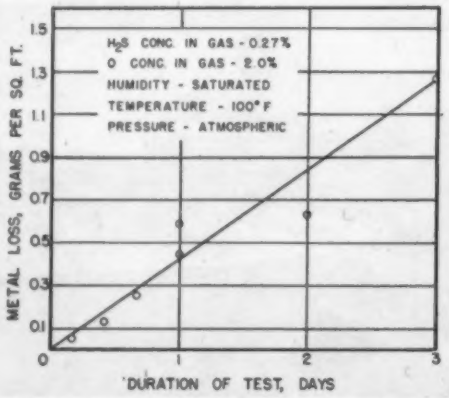


Figure 2. Progression of Gaseous H_2S Corrosion of Steel with Time.

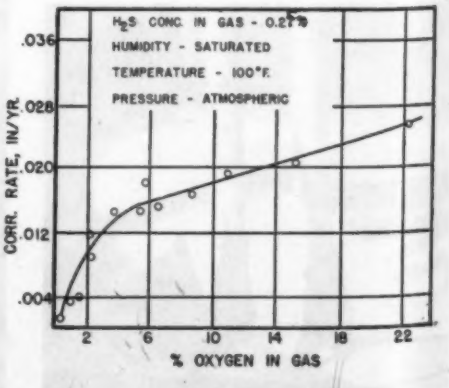


Figure 3. Effect of Oxygen on Gaseous H_2S Corrosion of Steel.

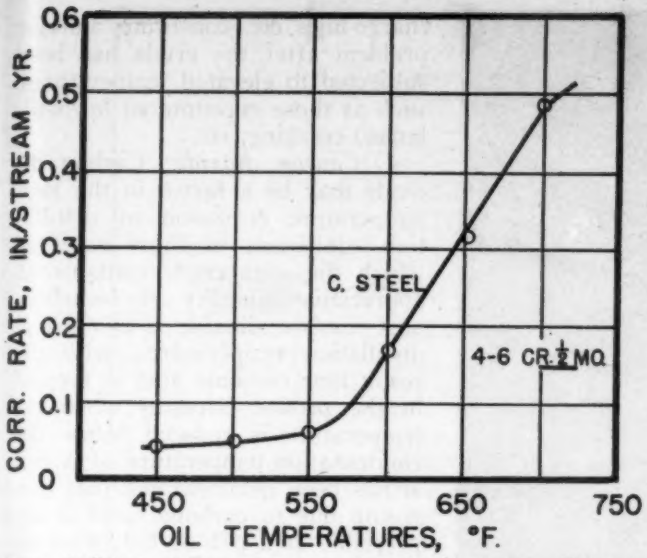


Figure 4. Effect of Temperature on Corrosion of Furnace Tubes in Processing Sour Crude from West Texas.

that occurs at temperatures between atmospheric and about 500° F. is usually classified as low-temperature corrosion. Corrosion that is experienced at temperatures above about 500° F. is usually classified as high-temperature corrosion. The presence of water is usually essential to low-temperature corrosion. Oxygen is another important factor in determining the extent to which refinery equipment is corroded under low-temperature conditions. Low-temperature corrosion is usually affected to an appreciable extent by the presence of inorganic salts in the water phase. The presence of water and inorganic salts may or may not be a factor in high-temperature corrosion. The presence of oxygen is not considered an important factor in high-temperature corrosion. Turbulence is an important factor in both low- and high-temperature corrosion.

Inasmuch as the electromechanical theory is generally accepted in explanation of the mechanism of all types of corrosion, no discussion is given of the theory of the corrosion involved in processing sour crudes in this paper.

The low-temperature corrosion encountered in processing sour crudes is primarily due to hydrogen sulfide, hydrochloric acid, water, oxygen, and carbon dioxide. The presence of water is considered necessary for any appreciable corrosion to occur in processing sour crudes at low temperature. A brief discussion of the other corrosive media involved in low-temperature corrosion of equipment involved in processing sour crudes is presented hereinafter.

1) Hydrogen-sulfide corrosion: The severe corrosion of equipment that is caused by hydrogen sulfide in sour crudes has been discussed in detail by several investigators. Corrosion by hydrogen sulfide at low temperature presents a somewhat different problem from the high-temperature hydrogen-sulfide corrosion that is encountered in certain refining processes, such as cracking. It is generally recognized that both oxygen and water are accelerators of low-temperature hydrogen-sulfide corrosion, and that both must be present before appreciable corrosion can occur. Figure 1 shows the effect of the concentration of hydrogen sulfide on the low-temperature corrosion of steel in an atmosphere of oxygen, natural gas, and water vapor.[23] It is indicated that, for a given concentration of oxygen, viz., 2 percent, the corrosion tends to be proportional to the hydrogen-sulfide content of the gas up to a concentration of approximately 0.10 percent. As the hydrogen-sulfide concentration of the gas is increased above about 0.10 percent, the corrosion rate decreases with an increase in hydrogen-sulfide concentration. This is an indication that the ratio of the oxygen concentration to the hydrogen-sulfide concentration is the controlling factor in the amount of corrosion experienced. Figure 2 shows how low-temperature hydrogen-sulfide corrosion of steel progresses with time.[23] These data indicate that the scale formed on the steel is not protective.

2) Dissolved oxygen: The presence of oxygen is

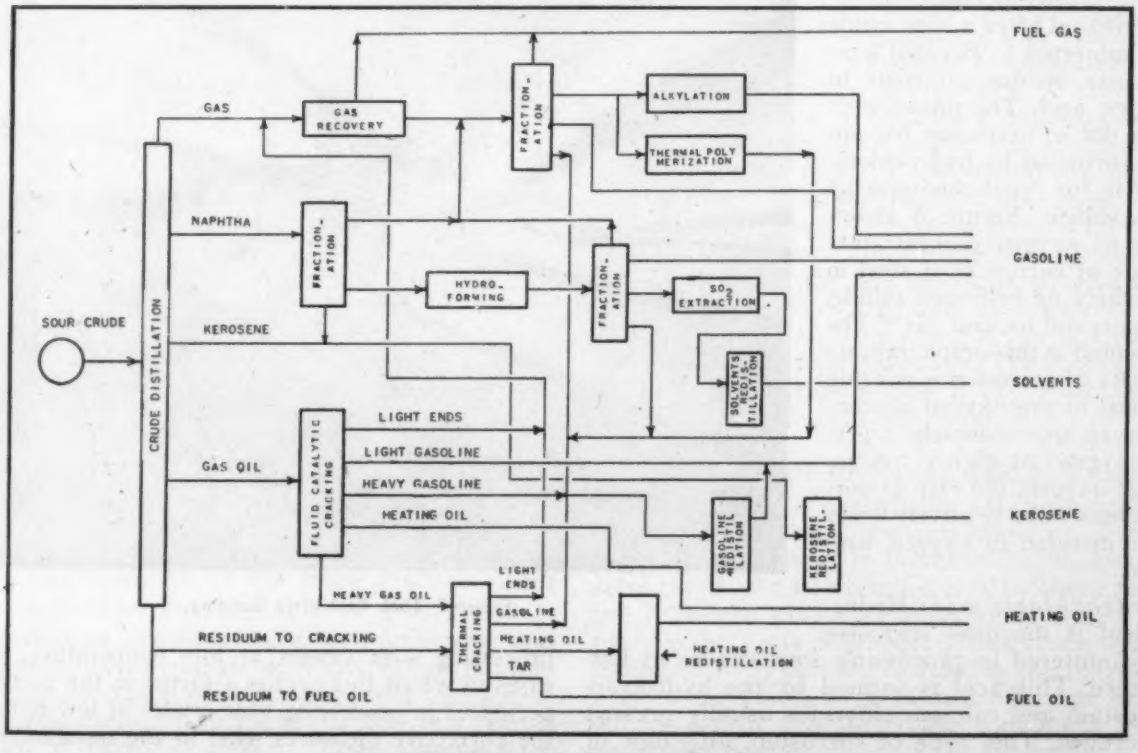


Figure 5. Flow Diagram for Processing Sour Crudes.

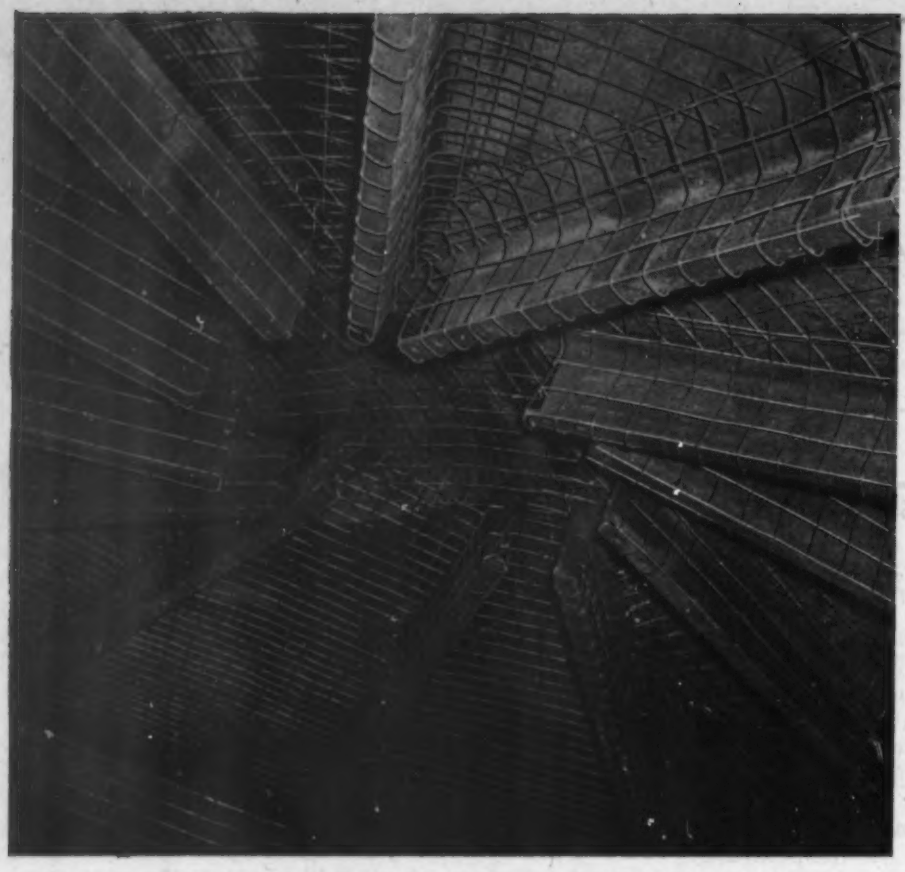


Figure 6. Tank Roof Showing Reinforcement before Guniting.

necessary for any appreciable amount of corrosion to occur in processing sour crudes under low-temperature conditions. This applies especially to the corrosion experienced prior to the distillation step, or to other processes involving appreciable rise in temperature of the crude. The low-temperature corrosion experienced after a sour crude has been subjected to elevated temperature may be due primarily to hydrochloric acid. The presence of oxygen is not so necessary for appreciable corrosion by hydrochloric acid as it is for rapid corrosion by hydrogen sulfide. Figure 3 shows the effect of oxygen concentration on the rate of corrosion of steel in an atmosphere of hydrogen sulfide, water vapor, and natural gas.[23] The data presented in this graph indicate that the rate of corrosion is roughly proportional to the oxygen concentration up to approximately 5 percent of oxygen. At higher concentrations of oxygen, the rate of corrosion increases at a relatively lower rate with increase in oxygen concentration.

3) Hydrochloric acid: Hydrochloric acid is the most corrosive agent encountered in processing sour crudes at low temperature. This acid is formed by the hydrolysis of magnesium and calcium chlorides usually present in sour crudes. This type of corrosion, although of no particular significance in sour-crude storage tanks, charge lines, etc., constitutes a major problem after the crude has been subjected to elevated temperatures, such as those encountered in distillation, cracking, etc.

4) Carbon dioxide: Carbon dioxide may be a factor in the low-temperature corrosion of distillation equipment in those cases in which the sour crude contains an appreciable quantity of bicarbonates. Carbon dioxide is evolved at distillation temperatures, with the result that carbonic acid is formed in the product streams when the temperature is reduced below the condensation temperature of water. It has been demonstrated that corrosion due to carbonic acid is appreciable at a pH of 6.0; whereas corrosion by the stronger acid, such as hydrochloric, is not particularly rapid at a pH above about 4.5. This is explained on the basis that total acidity rather than pH value determine the extent to which an acid corrodes iron. Because carbonic acid is a weak acid, its pH value is higher than that of a stronger acid of equal concentration. Carbonic-acid corrosion is usually a minor factor in the corrosion involved in processing sour crudes at low temperature. In those cases in which this acid is a factor in the corrosion experienced in processing sour crudes at low temperature, the corrective measures used in the mitigation of corrosion by the stronger acids are usually applicable.

Figure 7. Tank Roof after Guniting.

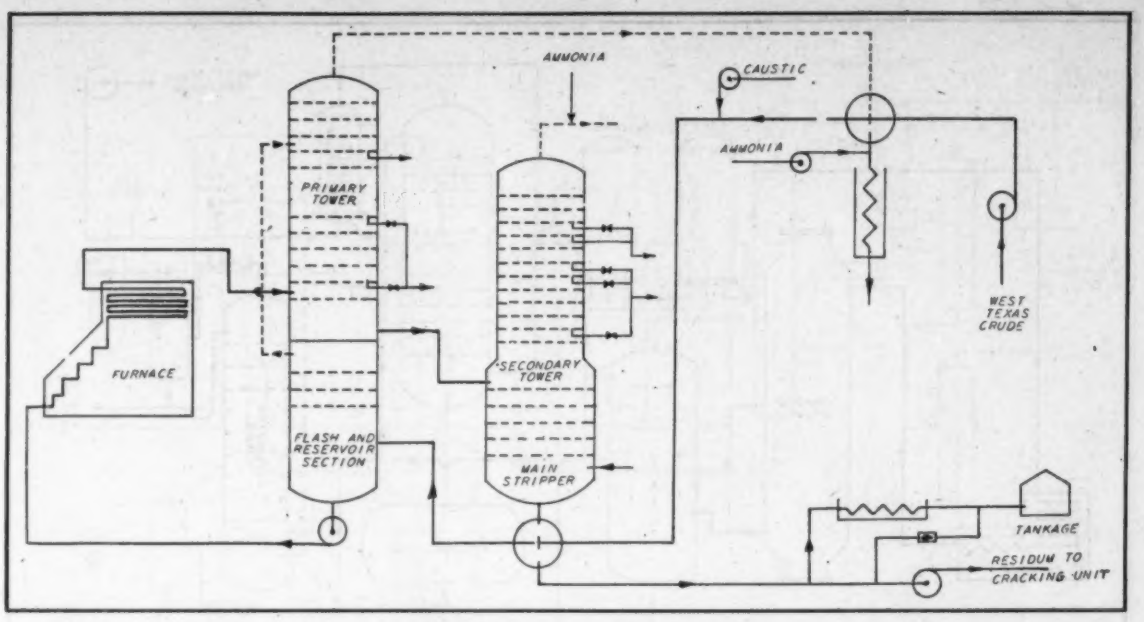


Figure 8. Flow Diagram of Pipe Still No. 1.

The corrosion which occurs in processing sour crudes at temperatures above about 500° F. is classified as high-temperature corrosion in this paper. Corrosion of this type is primarily due to sulfur compounds present in the crude oil. The presence of inorganic salts, particularly chlorides of calcium and magnesium, may have a marked effect upon the rate of corrosion experienced. The data presented in Table 1

TABLE 1

Effect of Hydrogen Chloride on Corrosion of 18-8 Alloy in Reforming Normal Hexane at 1320° F.

Charge to Unit	Corrosion Rate (Inches Per Year)
nHexane	None
nHexane plus hydrogen-chloride gas	0.1189

show the effect of hydrogen chloride on the rate of corrosion of 18-8 alloy in reforming normal hexane at a temperature of 1320° F. It will be observed that the corrosion rate was increased from zero to an average corrosion rate of 0.1189 inch per year by injecting anhydrous hydrogen chloride into the naphtha charge to the cracking unit. Of the various sulfur compounds present in sour crudes, it is generally agreed that free sulfur, hydrogen sulfide, and mercaptans are the most corrosive under high-temperature conditions. The corrosion that occurs as a result of the presence of other sulfur compounds in sour crudes is probably due to the action of the products formed by the decomposition of the relatively noncorrosive sulfur compounds at elevated temperatures. The data presented in Figure 4 show the effect of temperature on the rate of corrosion of steel in processing sour crudes.[16] Temperature and turbulence are probably the most important factors in corrosion by sulfur compounds at elevated temperature, other than the concentration of the more reactive sulfur compounds themselves.

The results of previous work have indicated that naphthenic acids may be an important factor in high-temperature corrosion of steel equipment.[2] As an example, it was found that naphthenic acid in a refined oil at 650° F. corroded steel at a rate varying from 0.06 inch per year to 0.2 inch per year, depending primarily upon the amount of acid present.

Flow of Sour Crude Through Refinery

A simplified flow diagram showing the flow of sour crude through the refinery is presented in Figure 5. It will be observed that the flow of a sour crude through a refinery constitutes an extremely complicated pattern. Corrosion is experienced throughout all the equipment in which the sour crude and its component parts are processed. A detailed discussion of all the corrosion experienced in processing sour crudes would be a major undertaking. For this reason, the present paper is limited to a discussion of some of the major corrosion problems encountered in storing, transporting, distilling, and cracking sour crudes and sour distillates.

Tanks

The corrosion of steel tanks in sour-crude and sour-distillate services is a problem of considerable significance in the operation of a refinery.[24] The corrosion is due to the action of hydrogen sulfide, oxygen and water. This type of corrosion was discussed in the first part of this paper. The roof and two top rings of the shell usually suffer the most corrosion. The corrosion of the bottoms of the tanks present the next most important problem. The corrosion of the shells of the tank, other than the two top rings, presents the least problem. The roofs of tanks in sour-crude service have a life varying from about 5.3 to 11.6 years, with 8.7 years being the average life. The roofs of tanks in sour-distillate service have a life from 3.8 to 14.3 years, with an average life of 7.9 years.

Means of Increasing Life of Tanks

1) Heavier steel construction: It is indicated that one of the most effective low-cost methods of increasing the life of a tank is to increase the thickness of the steel plates. As an example, it is common practice to use 3/16-inch plates for tank roofs. Based upon replacing the roofs when the plates reach a minimum thickness of 1/16-inch, increasing the original thickness of the plates by 1/16-inch offers an increase of 50 percent in the life of the roof for a 33-percent additional cost of material for the roof, or approxi-

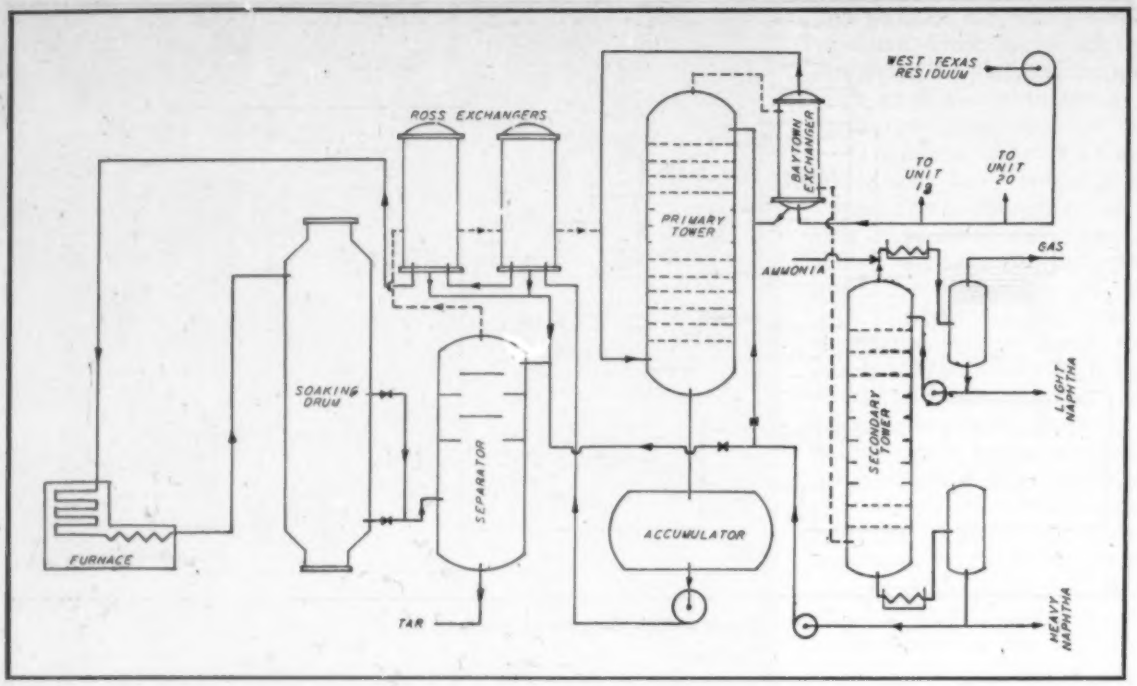


Figure 9. Flow Diagram of Residuum Cracking Unit.

mately a 20-percent additional over-all cost for material and labor; because, due to the longer life expected from the thicker plates, the yearly cost for installation and dismantling would, of course, be considerably reduced. As a result of the longer relative life of the rafters and other structural supports as compared with that of a roof of normal thickness, the use of thicker roof plates would more nearly equalize the replacements of these units of equipment. The use of 5/16-inch roof material could probably be justified for use on rundown tanks and tanks of intermediate size in sour service where the previous roof lasted less than about 6 years; whereas ¼-inch material would probably be more practical on larger tanks, where increased thickness of supporting members would be a factor.

2) Roofs of aluminum and galvanized steel: Twenty-seven aluminum roofs have been installed on rundown tanks in sour-distillate service at Baytown. The first of these roofs was installed in March, 1928. All of the others were installed between February, 1931, and February, 1936. No corrosion has been experienced on these roofs proper, other than the formation of a thin white coating, probably of aluminum oxide. Some galvanic corrosion has been experienced at the point where the roofs are riveted to the angle irons connecting the roofs and shells of the tanks. The same type of corrosion has been experienced where the roofs contact other steel supporting members.

In December, 1930, two rundown tanks were equipped with galvanized roofs. These roofs are still in good condition. The only evidence of corrosion is a white powdery material formed on the surface of the galvanized coatings.

The present practice is to install steel roofs on rundown tanks, rather than aluminum or galvanized roofs; because the average life of steel roofs is approximately 8.2 years, and an aluminum or galvanized roof, either of which costs approximately 2 to 3 times as much, would have to give at least 16 years of uninterrupted service to pay out the added investment. Further, steel roofs can be protected at a relatively low cost. This will be discussed later under the application of gunite coatings to tank roofs and shells.

3) Caustic washing: Caustic washing of distillates from sour crudes has effected an increase of 5.4 years in the average life of tank roofs in rundown and intermediate storage service. Maximum use is made of this method of increasing the life of tank roofs, because it is eventually necessary to remove hydrogen sulfide from sour distillates.

4) Flat roofs in crude service:[27] It has been observed that the roof and top rings of tanks in sour-crude service corrode more rapidly than do the shells. This difference in corrosion rate is attributed to some extent to the protective film of oil left on the shells when the tanks are emptied after having been filled. In some cases flat roofs are installed on tanks in crude service with provisions for filling the tanks and wetting the underside of the roofs once every 30 days, thereby maintaining a protective film of oil on the roof plates. This procedure has the advantage of reducing the vapor space of the tanks to a minimum, which is reported to be effective in reducing the corrosion of the roof plates.

Because of the apparent advantage to be derived from wetting the underside of the roof plates with crude oil, spray systems have been installed in some tanks whereby crude oil may be sprayed at regular intervals on the roof and the top shell plates, thereby maintaining a protective film of oil on these portions of the tanks.

It is indicated in some cases that the practice of wetting the roofs and shells of tanks with crude oil has effected an increase in the life of these portions of the tanks, but it is not conclusive at present that this type of protection is economically feasible for general application. This is due primarily to: 1) the high cost of the equipment required to spray or wet the roofs and other portions of tanks; 2) the difficulties experienced in dislodging the water film from horizontal surfaces; and, 3) interruptions in pumping schedules in order to wet the roofs at intervals of 3 to

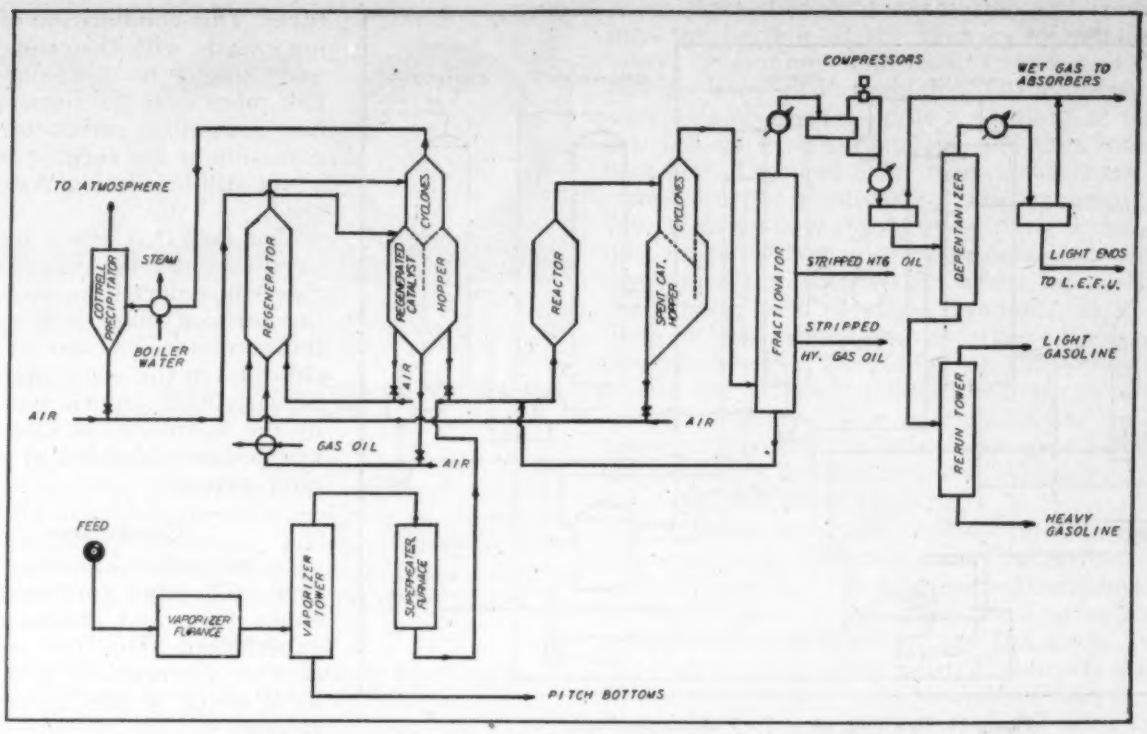


Figure 10. Simplified Flow Diagram of Fluid Catalyst Cracking Unit.

15 days, the frequency required to obtain any appreciable protection.

5) Protective coatings: A wide variety of nonmetallic materials has been tested by application to sand-blasted steel panels and exposure to sour vapors in crude and rundown tanks. Direct application of those materials showing the best protection were made to the roofs of tanks in sour service. None of these materials has given satisfactory protection for as long as two years. It is concluded from this work that none of the protective coatings manufactured at present is sufficiently resistant to justify its application to the roofs of tanks in sour-crude and sour-distillate services. Work on this phase of the problem is being continued by the industry in cooperation with the manufacturers of protective coatings, with the hope that a coating may be developed that will give satisfactory protection in sour-crude and sour-distillate services.

6) Gunite linings: Tank bottoms in sour-distillate service have been protected with concrete for about 18 years at the Baytown refinery. The first applications of concrete were made to the tank bottoms themselves, but later applications included a gunite lining extending about 18 inches up the shells of the tanks. Because of the good results obtained with coating of this type, gunite coatings were applied to the entire shells, roofs, and structural members of several tanks. The results obtained were satisfactory, and it is now common practice to gunite the inside surfaces of essentially all tanks in rundown and intermediate sour-distillate service.

The guniting of a tank is accomplished in three separate operations. These consist of the preparation of the surface, installation of the reinforcement, and application of the gunite. The final cleaning of a tank in preparation for guniting consists of sand-blasting to bare metal. It is common practice to sand-blast new tanks to remove the mill scale.

The preparation of a tank for guniting, the installation of the reinforcing, and the application of the gunite coating have been discussed in detail in the literature.[25] Figure 6 shows a tank roof with the reinforcement in place. Figure 7 shows a finished coating of gunite on a tank roof.

The cost of applying a gunite coating is in the order of 50 to 60 percent of the cost of a new tank. The application of gunite coatings to a tank is justified on the basis that they at least double the life of the tank. Normally, it is expected that gunite linings will more than double the life of a tank. The maximum life of such coatings in sour-crude and some distillate services has not been established.

Lines

It is not general practice to protect lines in sour-crude and distillate services, although it is recognized that they are subject to all the corrosive conditions that usually affect tanks in similar services, except that possibly less air is present in the lines. However, in some isolated cases ammonia or caustic solution may be injected into distillate lines to reduce corrosion. In those cases the corrosion is usually due to an inorganic acid, such as hydrochloric acid, rather than to compounds of sulfur.

Distillation Equipment

The data presented in this paper on the corrosion encountered in fractionating sour crude are based on the experience obtained in reducing West Texas crude to about 40 percent bottoms in a pipe still. The corrosion experienced in this particular operation parallels that experienced in other distillation operations in which sour crude constitutes the charge. It is believed, therefore, that a discussion of the corrosion experienced in this particular operation will suffice for the corrosion experienced in the distillation of sour crudes in general.

A simplified flow diagram of the pipe still used in reducing West Texas crude is shown in Figure 8. Briefly, the charge of West Texas crude flows

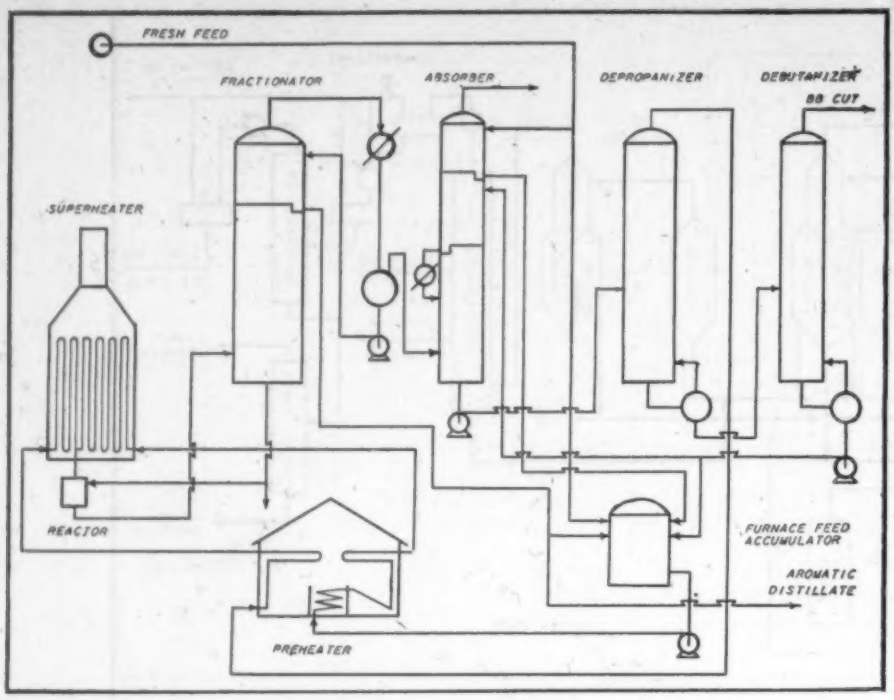


Figure 11. Flow Diagram of T.V.P. Butadiene Plant.

through a steam preheater, through naphtha vapor-to-crude and bottoms-to-crude exchangers, and is then flashed in the flash and reservoir section of the primary tower; the stabilized crude flows into the furnace. The stabilized crude is heated to a temperature of about 575° F., and flows from the furnace into the primary tower, where light naphtha is distilled overhead. The bottoms from the primary tower enter the secondary tower, where the remainder of the upper 60 percent of the crude is removed as overhead and side streams. The bottoms from the secondary tower are pumped through the bottoms-to-crude exchangers and then into a reduced-crude cracking unit.

In the early operation of this pipe still, rapid corrosion was experienced in the furnace tubes and overhead lines and condensers. Further details of this corrosion and the methods employed in its mitigation are presented in the following discussion.

Furnace Tubes

All the original tubes in the furnace of this pipe still were of low-carbon steel. Rapid corrosion of these tubes was experienced, necessitating replacements after 4 to 6 months. The first step in reducing this corrosion consisted of introducing caustic with the crude charge. It was found that caustic solution equivalent to 6 pounds of sodium hydroxide per 1000 barrels of crude oil could be injected with the charge without affecting the operation of the pipe still or of the residuum cracking units. This quantity of caustic reduced the corrosion of the furnace tubes, but replacements of tubes near the coil outlet was necessary after 6 to 8 months. Larger amounts of caustic progressively reduced furnace-tube corrosion at the pipe still, but limited operations at the residuum cracking units because of the rise in tube-metal temperature of the cracking-coil furnace tubes. As a result of this, caustic solution equivalent to 6 pounds per 1000 barrels of charge was injected with the crude charge, and 20 steel tubes at the coil outlet were replaced with 4 to 6-percent-chromium-0.5-percent-molybdenum tubes. This combination of injecting caustic with the crude charge and using 4 to 6-percent-chromium tubes near the furnace outlet has controlled satisfactorily the corrosion of the furnace tubes on a pipe still processing West Texas crude.

The fact that the injection of caustic with the charge to the pipe still reduced the corrosion of the furnace tubes indicates that the corrosion was due in part to chlorides in the crude charge, particularly hydrochloric acid formed by the hydrolysis of calcium and magnesium chlorides at elevated temperature.

Condensers

In the initial operation of this pipe still, rapid corrosion of the vapor lines and condensers was experienced. This corrosion was due to hydrochloric acid formed as a result of the hydrolysis of calcium and magnesium chlorides present in the crude charge. The injection of caustic with the crude charge effected a marked reduction in the amount of hydrachloric acid evolved, but did not eliminate it. The vapor lines are of low-carbon steel, the naphtha-vapor-to-crude exchanger contains admiralty-metal tubes in a steel shell, and the final condensers and coolers are cast iron of the coil-in-box type.

Ammonia is injected into the vapor lines upstream from the coil-in-box condensers to control the corrosion of this equipment. Ammonia injection and its control have been discussed on numerous occasions. In this particular case, the pH of the water condensed with the naphtha is maintained at 6.5 to 7.0. Water is injected upstream from the ammonia-injection point to prevent the deposition of ammonium salts in the vapor lines and condensers. Higher pH's are avoided to conserve ammonia. If higher pH's were maintained considerable quantities of the ammonia would be consumed in neutralizing hydrogen sulfide. This is of no particular significance in corrosion control under conditions of this type.

Frequently, ammonia is injected upstream from exchangers that are equipped with copper-bearing alloy tubes. It is normally assumed that copper-bearing alloys in condensers on sour-distillate streams would be corroded rapidly by ammonia when this neutralizer is injected upstream from such condensers, particularly if the pH of the water condensed with the sour distillate is maintained above the neutral point. It has been found in practice, however, that this is not the case. Several observations have been cited in an attempt to explain this apparent anomaly, but no satisfactory explanation has appeared in the literature. The results of some recent work on this phase of the problem indicate that certain sulfur compounds, such as hydrogen sulfide, mercaptans, carbon disulfide, etc., are very effective inhibitors for controlling the corrosion of copper-bearing alloys in atmospheres containing ammonia and oxygen. The results presented in Table 2 show the extent to which sulfur compounds reduce the corrosion of admiralty metal in a 10-percent-by-weight solution of am-

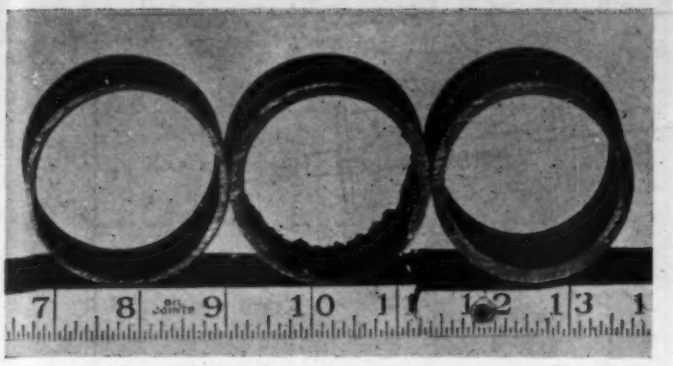


Figure 12. Examples of Uniformly Thinned, Pitted and New 18-8 Alloy Superheater Furnace Tubes.

monium hydroxide at room temperature for 48 hours.

In another case, admiralty metal under stress failed as a result of stress-corrosion cracking in an atmosphere containing ammonia, moisture, and oxygen after 3 hours of exposure. This test was repeated with 3.0 percent by weight of butyl mercaptan added to

TABLE 2

Effect of Sulfur Compounds on Corrosion of Admiralty by 10 Percent Ammonium Hydroxide Solution

Sulfur Compound Added	Percent Added	Corrosion Rate (Inches Per Year)	Reduction of Corrosion Rate (Percent)
None	.....	0.5895	.....
Ammonium hydrosulfide	0.025	0.0368	93.8
Ammonium hydrosulfide	0.05	0.0187	96.8
Ammonium hydrosulfide	0.10	0.0320	94.5
Ammonium hydrosulfide	0.30	0.0096	98.4
Butyl mercaptan	0.25	0.3930	33.4
Butyl mercaptan	0.50	0.0678	88.5
Butyl mercaptan	2.0	0.0442	92.6
Butyl mercaptan	2.5	0.0315	94.5

the ammonium-hydroxide solution over which the stressed admiralty metal was suspended. After 194 hours of exposure, there was no evidence of failure as a result of stress-corrosion cracking. Equally good results were obtained with hydrogen sulfide and carbon disulfide.

Corrosion in other portions of the pipe-still equipment is not a problem of any consequence.

Cracking Units

In selecting material for presentation and discussion on the corrosion encountered and on the methods employed in its mitigation in cracking reduced sour crudes and sour distillates, the corrosion experienced in the operation of three separate types of cracking processes has been chosen. These processes include thermal cracking of reduced West Texas crude, cracking of sour distillate in a fluid catalyst cracking unit, and vapor-phase reforming of a light naphtha. The discussion of the corrosion experienced in these cracking processes is limited to some of the major corrosion problems involved, because it would be beyond the scope of this paper to discuss in detail all the corrosion experienced. It is believed that the corrosion discussed in connection with these three cracking processes is representative of the major corrosion problems encountered in cracking in general.

Thermal Cracking Unit

Reduced West Texas crude from a pipe-still operation is cracked in a partial-yield operation on a thermal cracking unit. Figure 9 shows a flow diagram of this type of cracking operation. The residuum charge flows through a heat exchanger and is discharged into the bottom of the primary distillation tower, where it combines with recycle stock, and then flows into the hot-feed accumulator. The total feed from the accumulator is pumped at a pressure of about 1200 psi through heat exchangers, and then into the furnace coil. The feed flows from the furnace coil at a temperature of about 900° F. and a pressure of about 700 psi down-flow through a soaking drum. The feed then flows through a pressure-reducing valve and is discharged into a separator, where tar is removed at the bottom and the light products pass overhead through heat exchangers, and then into the primary disillation tower. Recycle stock flows from the bottom of the primary distillation tower into the total feed accumulator. The overhead from the primary tower passes through heat exchangers and then into the secondary distillation tower, where it is separated into light and heavy naphtha.

1) Furnace tubes: The first 72 tubes in the convection section of the furnace are low-carbon steel, the next 30 tubes are 4 to 6-percent chromium steel, and the last 16 tubes in the convection section and the 39 tubes in the radiant section are 18-8 alloy. Originally all tubes in this furnace were of low-carbon steel, but those in the radiant section and the last 16 tubes in the convection section were replaced with 18-8 alloy because of the rapid corrosion experienced. It was later found necessary to replace the next 30 tubes in the convection section with 4 to 6-percent chromium-steel tubes. The life of steel tubes in the radiant section of the furnace was as low as 30 days, and the life of the last 16 tubes in the convection section was in the order of 3 months. In general, the performance of 18-8 alloy tubes in this type of cracking operation has been satisfactory. Many of the original tubes are still in service after more than 100,000 hours of operation. During this period some tubes have failed, whereas others have undergone some rather marked changes in their physical properties. It has been found that the tubes possess entirely different physical characteristics. Some are still ductile; others are brittle, but can have ductility restored by heat treatment; whereas others are permanently brittle. On the basis of certain examinations, it has been postulated that the deterioration and, possibly, the failure of the 18-8 alloy tubes are due to structural changes at the grain boundaries. These changes are believed to be progressive in nature, and are dependent upon time, temperature, and stress. It is further postulated that these structural changes are due to the decomposition of the austenite, which results in the precipitation and growth of highly alloyed ferrite. When these precipitated areas have reached a certain size, cracking occurs under certain conditions with respect to time, temperature, and stress. Once cracking has occurred, the tube is permanently brittle, but at any time prior to the cracking the tube is either ductile or can have ductility restored by proper heat treatment.

2) Hot-oil lines: In the initial operation of this plant, rather severe corrosion was experienced in all lines through which oil was pumped at a temperature above about 450 to 500° F. In order to eliminate this type of corrosion, essentially all of the lines downstream from the furnace have been replaced with 18-8 alloy lines. The installation of 18-8 alloy lines in hot-oil service has essentially eliminated corrosion in this service.

3) Soaking drum: The soaking drum is of low-carbon steel. The temperature conditions are such in this

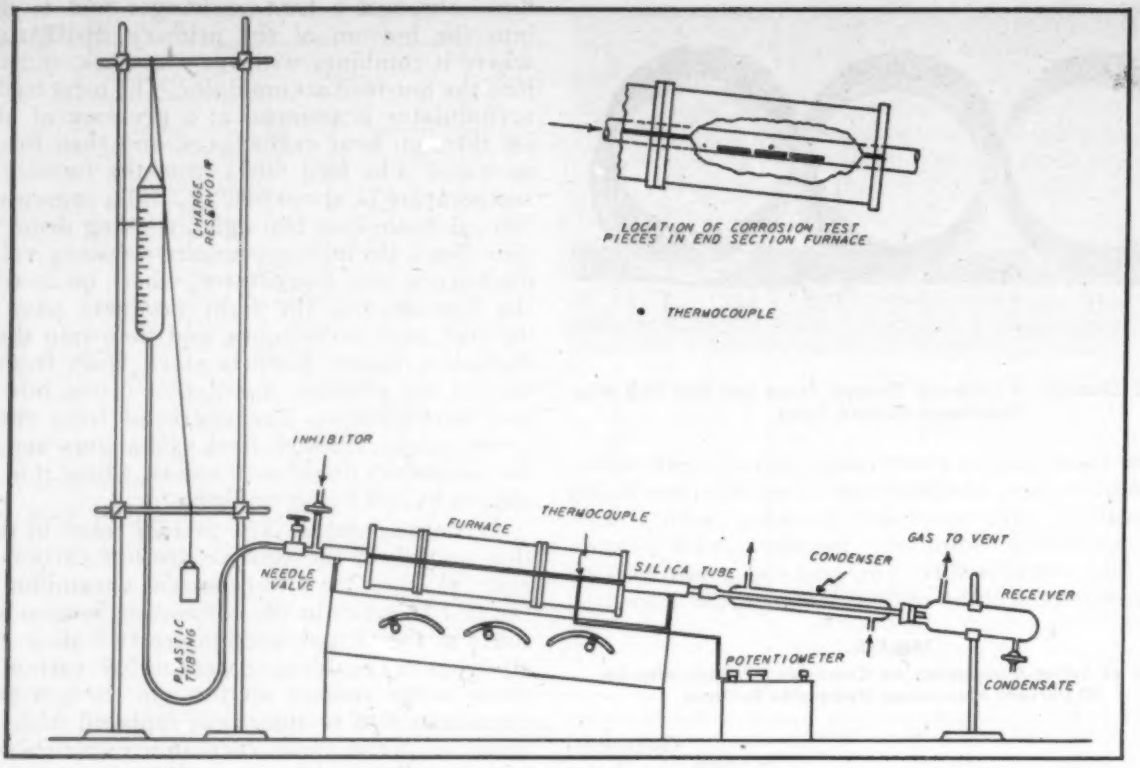


Figure 13. Apparatus Used in Corrosion Tests

drum that rapid corrosion would occur if the drum were not protected. The drum is protected against corrosion by the application of a reinforced layer of gunite about 3 inches in thickness. This type of protection is very effective in reducing corrosion of the drum, but the coating is frequently damaged in the process of removing coke from the drum. This necessitates frequent repairing and replacing of the gunite lining. This type of protection is so effective that these drums are in satisfactory operating condition after almost 20 years of service in cracking reduced West Texas crude.

4) Separator: The separator was originally of all-steel construction. Rapid corrosion was experienced in this piece of equipment, and it was necessary to apply protection to avoid failure of the separator. The protection employed consists of guniting the shell and installing an 18-8 alloy lining in manhole openings, etc. The plates in the separator are of steel, but are being replaced with 4 to 6-percent chromium steel as they fail. All supports are 18-8 alloy.

5) Bubble towers: The bubble towers are of low-carbon steel, and are equipped with steel plates and cast-iron bell caps, except for the three bottom plates in the primary bubble tower. These plates are of 11 to 13-percent chromium steel, with bell caps of the same material. The shell of the primary tower is protected with a gunite lining.

6) Accumulator: The accumulator is of all-steel construction, and is protected with a gunite lining.

7) Preheaters: The exchanger is the primary preheat circuit, and is of all-steel construction. Corrosion is not a problem in this preheater. The exchangers in the secondary preheat circuit are equipped with steel shells and 18-8 alloy tubes. The steel shell is protected by metallizing with aluminum. This protection is satisfactory and, as a result, maintenance costs are low.

8) Neutralizers: It was found that the injection of caustic, 6 pounds per 1000 barrels of crude, into the crude charge to the pipe still resulted in an appreciable reduction in the corrosion of the thermal cracking unit in which the residuum from the pipe still was processed. Larger quantities of caustic could be tolerated in the pipe still, but any increase in the amount of caustic injected into the charge to the pipe still, i.e., above 6 pounds per 1000 barrels of crude, materially affected the service factor on the thermal cracking unit as a result of the rapid rise of metal temperatures in the furnace tubes. It was found, however, that an increase in the caustic injected into the charge to the pipe still effected a corresponding decrease in the corrosion of the cracking equipment. An experimental study of lime injection into the cracking unit was made to determine whether the corrosion of cracking equipment (occurring while 6 pounds of caustic per 1000 barrels of crude was being injected into the charge to the pipe still) could be reduced further by injecting lime ahead of the cracking-unit furnace. It was found that, when 250 pounds of lime per 1000 barrels of residuum was injected, the life of the steel tubes in the convection section of the cracking-unit furnace was increased by about 40 percent, and the overall loss of iron from the unit was decreased by 20 to 30 percent. The injection of lime was never put into practice, because the increase in the sediment content of the fuel from the cracking operation was more than could be tolerated.

Ammonia is injected into the overhead vapor line from the secondary bubble tower to neutralize inorganic and organic acid and, thus, reduce the corrosion of the overhead vapor line and condensers as well as other equipment downstream from the cracking unit.

Corrosion, as such, is not a problem of any particular significance in the operation of fluid catalyst cracking units. Some difficulty is experienced as a result of erosion of certain portions of the equipment, but serious corrosion has not been experienced except in one case. This corrosion occurred in the 18-8 alloy

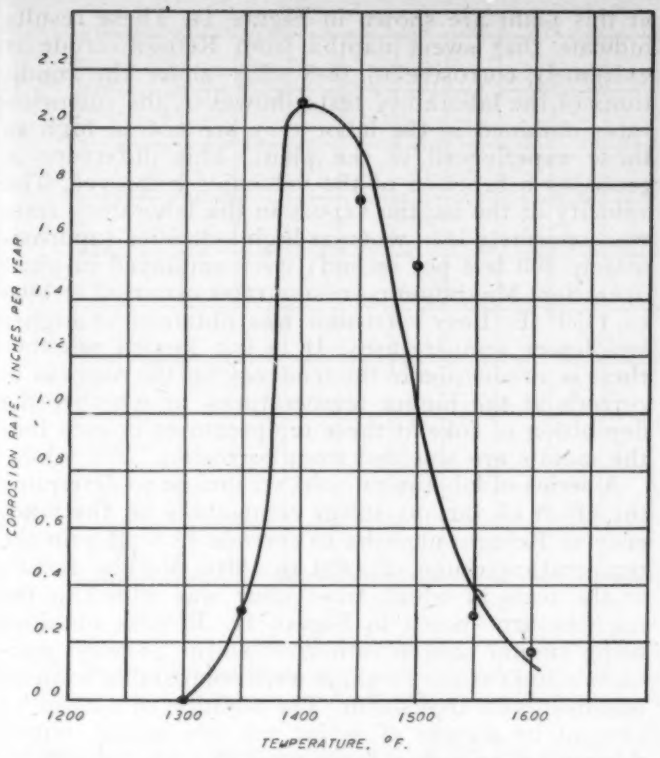


Figure 14. Corrosion Rates of 304-Type Stainless Steel by Refugio Naphtha.

vapor lines leading to and from the reactor on an upflow unit, and was intergranular in nature. A simplified flow diagram of this type of unit is shown in Figure 10. The vapor lines in which this stress-corrosion cracking occurred were lined at points of turbulence with carbon-steel wear plates to absorb erosion caused by the catalyst. The intergranular cracks that occurred in these vapor lines were located behind the carbon-steel wear plates. The extent of this type of corrosion was such that it made the unit essentially inoperable.

In the investigation of this corrosion problem, it developed that there was a small annular space between the wear plates and the 18-8 alloy vapor lines. Catalyst and hydrocarbon vapors circulated at low velocity in this small annular space when the unit was on stream. During this time, the carbon-steel wear plates were corroded rather rapidly by hydrogen sulfide at the operating temperature of about 950° F. As the velocities in the annular space between the wear plates and the 18-8 alloy vapor lines were low, a relatively thick scale of iron sulfide was formed on the carbon-steel wear plate.

It is well-known that 18-8 alloy loses some of its resistance to corrosion when heated for appreciable lengths of time at temperatures of 950 to 1400° F. This loss in resistance to corrosion is attributed to the precipitation of chromium carbide along the grain boundaries. This leaves the metal adjacent to the grain boundaries low in chromium content and, as a result, susceptible to corrosion when placed in the proper environment. When the environment is an acid solution of relatively low concentration, the attack is usually concentrated at the grain boundaries, and failure occurs as a result of intergranular corrosion, or stress-corrosion cracking.

In this particular case, it was postulated that, during the operation of the unit, beds of catalyst collected in the annular space between the wear plates and the 18-8 alloy vapor line. During the turnarounds, when the unit was steamed free of hydrocarbons, the beds of catalyst between the wear plates and 18-8 alloy vapor lines were wetted with water. After the unit had been opened, air was present for the oxidation of the iron sulfide to iron sulfite and sulfate. These salts hydrolyzed, forming sulfurous and sulfuric acid which attacked the 18-8 alloy—resulting in failure of the metal along the grain boundaries. In an attempt to duplicate this type of corrosion, stressed specimens of susceptible 18-8 alloy were imbedded in a mixture of catalyst and iron sulfide and wetted with water. After several days, the stressed 18-8 alloy failed as a result of intergranular corrosion.

As a result of this work, the carbon-steel wear plates were removed from all 18-8 alloy tubes and vessels in the catalyst system, and wear plates of 18-8 alloy were installed. This eliminated the formation of iron sulfide between the wear plates and the 18-8 alloy lines and, as a result, eliminated the intergranular corrosion of the 18-8 alloy lines and vessels.

Vapor-Phase Cracking[26]

This section of this paper is not concerned directly with corrosion experienced in processing sour crude. This discussion is included, however, to emphasize that all corrosion experienced in cracking operations at elevated temperatures is not necessarily due to sulfur or its compounds. It is shown in the discussion that follows that there are cases in which appreciable quantities of sulfur are extremely effective in reducing corrosion of 18-8 alloy furnace tubes in high-temperature cracking service.

During the recent war the true vapor-phase process was employed in reforming 200 to 400°-F.-boiling-range naphtha at a temperature of 1310 to 1325° F. for the production of butadiene. A flow diagram of the true-vapor phase process as it was employed in the production of butadiene is shown in Figure 11. Three plants of this type were constructed and put into operation. In the initial operation of one of these plants, a peculiar and acute type of corrosion was experienced which resulted in the failure of the 304-type 18-8 alloy tubes in the naptha circuit of the superheater furnace after only 120 hours of operation. Similar corrosion was experienced at the other two plants. One of these plants continued in operation, but the other was converted into a unit for the production of gasoline after several months of intermittent operation during which attempts to develop methods of suppressing the corrosion were unsuccessful. It is believed that the type and degree of corrosion experienced at these plants are without parallel.

In the initial stages of reforming sweet naptha from Refugio crude at one of these plants, the unit was rendered inoperable because of the rapid corrosion of the 18-8 alloy furnace tubes in the naptha circuit. An investigation of the corrosion experienced at this plant was undertaken at the request of the Petroleum Administration for War. Examination of a number of furnace tubes indicated that the failures were due primarily to uniform thinning of the tube wall. However, severe pitting did occur, particularly in the tubes near the inlets to the superheaters, and in new tubes in the outlet passes when they were first put into operation. Figure 12 shows examples of both the general thinning and the pitting type of corrosion experienced, and a section of a new tube. The new tubes were 2 ins. in diameter, with a wall

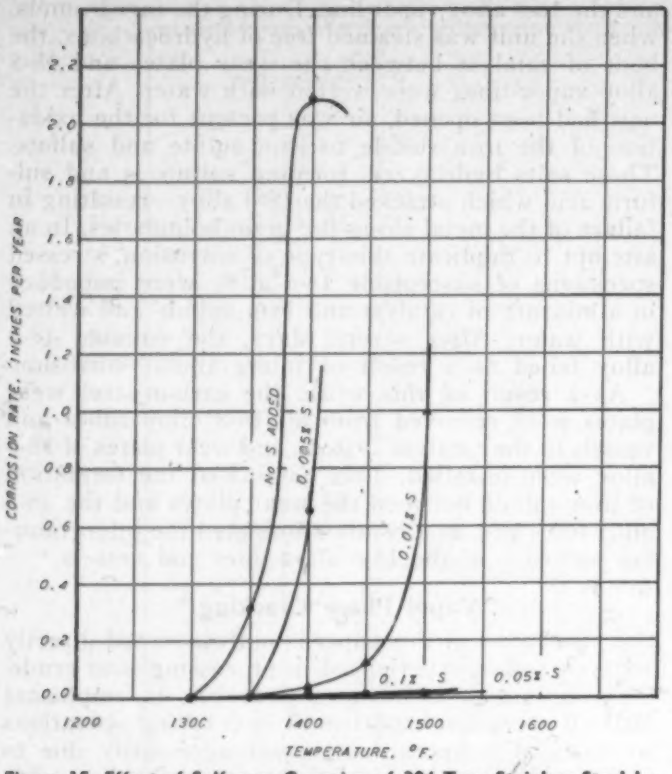


Figure 15. Effect of Sulfur on Corrosion of 304-Type Stainless Steel by Refugio Naphtha.

thickness of 0.156 ins. It will be observed that one side of the tube is severely pitted, whereas the other is only slightly attacked. The corrosion was limited primarily to the center sections of the tubes, and was more severe on the side (internal surface) next to the furnace. The end sections and return bonds were relatively free of corrosion. This indicates that metal temperature was an important factor in the corrosion.

Early in this investigation, extensive analytical data were obtained on the naptha charge and the recycle streams to the furnace, but none of these results indicated the cause of the corrosion.

Inasmuch as it was not possible to determine the cause of the corrosion experienced at this plant by analyzing the charge stocks and recycle streams, bench-scale laboratory studies were conducted (under conditions approximating those existing in the plant) for the purpose of developing the most feasible method of minimizing the corrosion.

The equipment employed in the laboratory investigation is shown in Figure 13. The various napthas were cracked in a quartz tube (0.875-inch inside diameter) at essentially atmospheric pressure, and at temperatures comparable to those employed in plant practice. A naptha charge rate of 200 mililitres per hour was used in these tests. The temperatures were measured with a thermocouple placed against the outside of the quartz tube in the center of the end section of the furnace. Three 18-8 alloy test specimens (cut from a furnace tube from the butadiene plant) were placed to end in the quartz tube at the center of the outlet section of the furnace. The location of the metal test specimens is shown in Figure 13. The test periods varied in length, but most of the runs were for 20 hours. The corrosion rates were calculated by the conventional weight-loss method.

The results obtained in tests that were made for the purpose of reproducing the corrosion experienced at this plant are shown in Figure 14. These results indicate that sweet naptha from Refugio crude is extremely corrosive to 18-8 alloy under the conditions of the laboratory tests; however, the corrosion rates obtained in the laboratory are not so high as those experienced in the plant. This difference is probably a function of the velocities employed. The velocity of the naptha vapors in the laboratory tests was extremely low, whereas high velocities (approximately 200 feet per second) were employed in plant operation. Maximum corrosion rates occurred at 1400 to 1450° F. Less corrosion was obtained at higher and lower temperatures. It is not known whether there is a reduction in the tendency for the naphtha to corrode at the higher temperatures, or whether the deposition of coke at these temperatures is such that the metals are shielded from corrosion.

A series of laboratory tests was made to determine the effect of various sulfur compounds on the tendency of Refugio naphtha to corrode 18-8 alloy in the temperature range of 1300 to 1500° F. The results of the tests in which free sulfur was added to the naphtha are shown in Figure 15. Results obtained using similar concentrations of sulfur as butyl mercaptan and carbon disulfide were comparable to those obtained with free sulfur. The addition of 0.05 to 0.5 percent by weight of sulfur (as free sulfur, butyle mercaptan, or carbon mercaptan, or carbon disulfide) to the naphtha essentially eliminates its tendency to corrode 18-8 alloy. It is indicated that lower amounts of sulfur effect marked reductions in the corrosion, but more uniformly consistent results were obtained on adding 0.5 percent by weight, and higher, of sulfur. It was observed also that the addition of sulfur essentially eliminated coke deposition on the 18-8 alloy. Both of these observations have been confirmed by the results obtained in plant operation.

The effect of the addition of small amounts of distilled water on the tendency of the naphtha to corrode 18-8 alloy under cracking conditions is shown in Figure 16. These data indicate that the corrosion can be eliminated by adding 0.06 to 0.4 percent by weight of water to the naphtha. Larger amounts of water are not detrimental, but are unnecessary. It was observed that the addition of water also effected a marked reduction in the tendency of the naphtha to coke 18-8 alloy under the test conditions employed. No plant experience is available to substantiate these results.

Further tests in the equipment previously described indicated that carbon dioxide, ethyl alcohol, acetic acid, hydrogen sulfide, and sulfur dioxide are satisfactory inhibitors for controlling the type of corrosion experienced at this butadiene plant. The optimum quantities of these materials required to maintain corrosion and coke deposition at a minimum were not established, because sulfur and water had been found to be adequate for this purpose.

Several exploratory corrosion tests were made in attempts to establish the cause of the corrosion. However, none of these tests gave an indication as to the causative agent. It was shown that the removal of active sulfur compounds, peroxides, chlorides, and inorganic salts (by refluxing with molten metallic sodium followed by washing with distilled water and filtering) effected no reduction in the tendency of the naphtha to corrode 18-8 alloy. The addition of such compounds as benzoyl peroxide, magnesium chloride, acetylene, *iso*butylene, normal butane, butadiene, furfural, aniline, etc., to a non-corrosive naphtha (normal

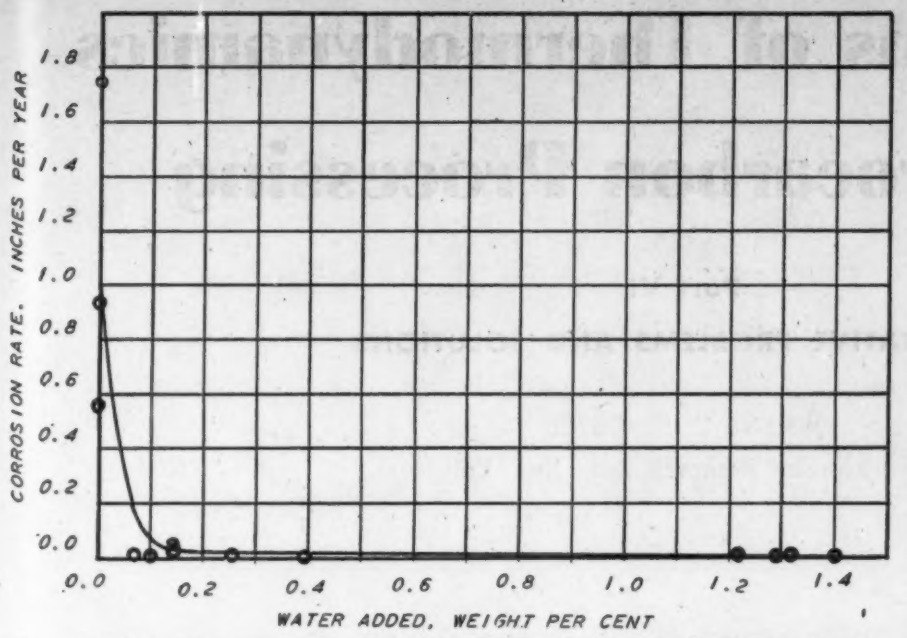


Figure 16. Effect of Distilled Water on Corrosion of 304-Type Stainless Steel by Refugio Naphtha at 1450° F.

hexane) failed to induce corrosion under the cracking conditions employed. Various types of essentially pure hydrocarbons (paraffins, aromatics, naphthenes, and unsaturates) were found non-corrosion.

The results of corrosion tests in which oxygen was excluded from the naphtha from Refugio crude indicate that dissolved oxygen is not a significant factor in the corrosion experienced at this butadiene plant.

In order to evaluate the possible effect of charge stock on the corrosion of the 18-8 alloy furnace tubes, a trial run was made at this plant employing a highly paraffinic naphtha (normal hexane) as fresh feed. This run was in progress for approximately 700 hours without any evidence of corrosion of the superheater furnace tubes. Samples of various naphtha streams were taken for analytical examination during this run. The results obtained on analyzing these samples were not materially different from those obtained on comparable samples when processing Refugio naphtha, and are not conclusive as to the cause of the corrosion when processing Refugio naphtha, or to the lack of corrosion when charging normal hexane.

When the trial run on normal hexane was terminated, a run was made on freshly distilled gas-blanketed Refugio naphtha to which sufficient butyl mercaptan had been added to maintain 0.005 to 0.006 percent by weight of sulfur (as butyl mercaptan) in the furnace charge. This run was terminated after 118 hours because of the failure of an outlet tube in a naphtha pass. The results obtained on inspecting the furnace tubes were not so discouraging as the short length of the run indicated. The seventh and eighth tubes were severely attacked, whereas only incipient corrosion had occurred on the sixth tube. In previous runs that were terminated in about the same length of time because of tube failures, it was found that the fifth, sixth, seventh and eighth tubes were corroded to about the same extent with only 0.01 to 0.02 inch of the original 0.156-inch wall thickness remaining.

Because of the results obtained in the first run employing sulfur as an inhibitor, and because the laboratory results indicated that sulfur in concentrations up to 0.5 percent by weight was beneficial rather than harmful to the 18-8 alloy, a second run was made employing a minimum of 0.05 percent of added sulfur based on the furnace charge. Inspection of the furnace tubes after 135 hours of operation showed that no corrosion had occurred in any of the tubes. As a result of this successful run, operation of the unit was continued. Carbon disulfide was injected in a quantity sufficient to maintain approximately 0.3 percent by weight of sulfur (as carbon disulfide) in the furnace feed. The unit was in operation for over 10,000 hours without any evidence of corrosion in the naphtha tubes of the superheater furnaces. After 2764 hours of operation with sulfur injection, a failure occurred in one of the outlet tubes of the recycle-gas circuit. On inspection, the failure was found to be due to pitting at random in the gas tube. Only one other outlet gas tube was found to be pitted sufficiently to require replacement. None of the other tubes in the recycle gas circuit was attacked. For several weeks prior to the failure of the gas tube, the concentration of hydrogen sulfide in the recycle gas stream was subject to considerable variation. A large quantity of absorber overhead gas has been used for recycle gas, which resulted in a substanially lower concentration of hydrogen sulfide in the recycle-gas stream. It is a contributing factor in the corrosion of the gas tubes.

The addition of sulfur to the furnace feed essentially eliminated coke deposition in the furnace tubes. With added sulfur in the charge, only a very thin, hard scale formed on the inside of the tubes. Analysis of this scale showed it to be essentially carbon.

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Applications of Thermodynamics To Hydrocarbon Processing

Part VI

ILLUSTRATIVE PROBLEMS AND SOLUTIONS

WAYNE C. EDMISTER
Hydrocarbon Research, Inc., New York

PRACTICALLY all hydrocarbon-processing operations involve energy changes and equilibria, for which useful calculation tools and methods are developed by applying thermodynamic functions to hydrocarbon properties. In the preceding chapters the fundamentals and formulations of thermodynamics were developed and applied to perfect gas and non-hydrocarbon fluids. In order that these concepts be fully understood and appreciated before starting on the complex subject of hydrocarbon thermodynamics, this chapter will be devoted to various problems and their solutions. These examples are intended to illustrate the engineering applications of the first and second laws of thermodynamics. The problems included are of the classical type, dealing with perfect gas, air, steam, etc. An understanding of these applications prepares the way for hydrocarbons.

THE examples presented in this chapter were selected to illustrate the applications of thermodynamics in making almost every type of energy calculations. The problems are simple and the solutions explained in detail. Miscellaneous non-hydrocarbon fluids and materials are used. Some of the solutions involve using equations already derived and some go back to the basic concepts and laws.

These examples are not only intended to illustrate the fundamentals but in addition to demonstrate the practical and very important problem of conversion from work to heat units in a consistent manner when making the numerical calculations involved in applying thermodynamics to physically real operations. The latter objective is more important than the first.

These examples come from courses the author took, from courses the author taught, from problems encountered in industrial work and some are original in this article.

Example 1

Problem: A perfect monatomic gas for which $C_v = \frac{3}{2}$ R and $C_p = \frac{5}{2}$ R is being changed continuously from a pressure of 1 atm. and 273.1° K to a pressure of 0.5 atm. and a temperature of 546.2° K.

Calculate the work done by the gas, the heat absorbed, ΔE, ΔH, and ΔS for each of the following paths:

(Path 1) a. Expand isothermally and reversibly to the final pressure.
b. Heat at constant pressure to the final temperature.

(Path 2) a. Expand isothermally to the final volume.
b. Heat at constant volume to the final temperature.

(Path 3) a. Expand adiabatically to the final pressure.
b. Heat at constant pressure to the final temperature.

(Path 4) a. Expand adiabatically to final volume.
b. Heat at constant volume to the final temperature.

The above changes are assumed reversible.

Solution: This problem will be solved in engineering units for one pound-mol. of the perfect gas assuming the changes occur in a flow process.

$P_1 = 14.7$ psia $\quad P_2 = 7.35$ psia

$T_1 = 492°$ R $\quad T_2 = 984°$ R

$V_1 = 359$ cu. ft. $\quad V_2 = 359 \frac{984}{492} \frac{14.7}{7.35} = 1436$ cu. ft.

The intermediate conditions for each path will be calculated by the following equations, which are based on equations 115 and 117 and the given values of C_p and C_v.

$$TV^{2/3} = \text{const.} \quad (a)$$

$$PV^{5/3} = \text{const.} \quad (b)$$

From equation (a)

$$T_2 = T_1\left(\frac{V_1}{V_2}\right)^{2/3} \text{ and} \quad (c)$$

$$V_2 = V_1\left(\frac{T_1}{T_2}\right)^{3/2} \quad (d)$$

From equation (b)

$$P_2 = P_1\left(\frac{V_1}{V_2}\right)^{5/3} \quad (e)$$

$$V_2 = V_1\left(\frac{P_1}{P_2}\right)^{3/5} \quad (f)$$

Combining equations (d) and (f) gives

$$T_2 = T_1 \left(\frac{P_2}{P_1}\right)^{2/5} \qquad (g)$$

The above equations are used in calculating the intermediate conditions for each of the four paths. The calculations and the results for the intermediate conditions are given in the following tabulation:

PATH	Pressure	Temperature	Volume
1.....	7.35	492	$359\left(\frac{14.7}{7.35}\right)=718$
2.....	$14.7\left(\frac{359}{1436}\right)=3.67$	492	1436
3.....	7.35	$492\left(\frac{7.35}{14.7}\right)^{2/5}=373$	$359\left(\frac{14.7}{7.35}\right)^{3/5}=544$
4.....	$14.7\left(\frac{319}{1436}\right)^{5/3}=1.46$	$92\left(\frac{359}{1436}\right)^{2/3}=195$	1436

The temperature-volume plot given as Figure 12 shows the four paths, each of which is divided into two parts. It will be noted that path 1a coincides with part of path 2a, and that this occurs also for the other paths. Reference to Figure 12 will help make the following heat, work, and entropy calculations more clear.

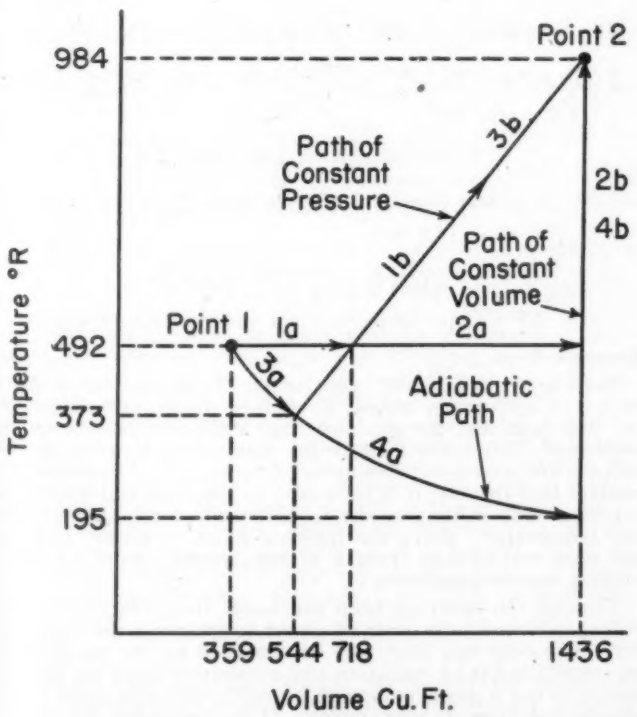

FIGURE 12

Alternate Paths for Change of Conditions for Perfect Gas

Heat and work calculations will be made for each path.

Path 1

(a) Isothermal expansion to final pressure
ΔE & $\Delta H = 0$ for perfect gas and

$$Q = W = -\int_{P_1}^{P_2} V dP = -\int_{P_1}^{P_2} RT/P \, dP = -\int_{P_1}^{P_2} RT \, d \ln P$$

$Q = W = (1.99)\ (492)\ (2.3)\ (\log_{10} 2) = 676$ Btu/mol.

$$\Delta S = \frac{676}{492} = 1.38 \text{ Btu/mol} - °R.$$

(b) Isobaric heating to final volume

$W = P\Delta V = R\Delta T = (1.99)\ (984\text{-}492) = 980$ Btu/mol

$Q = \Delta H = \int C_p dT = 5/2\ R\ (984\text{-}492) = 2460$ Btu/mol

$\Delta E = \Delta H - \Delta PV = \Delta H - R\Delta T = 2460 - (1.99)\ (492)$
$= 1480$ Btu/mol

$$\Delta S = \int \frac{C_p}{T} dT = 5/2R \frac{dT}{T} \int = 5/2\ (1.99)\ (2.3) \left(\log_{10} \frac{984}{492}\right) = 3.46 \text{ Btu/mol-°R.}$$

Path 2

(a) Isothermal to final volume
ΔE and $\Delta H = 0$

$$Q = W = (1.99)\ (492)\ (2.3) \left(\log_{10} \frac{14.7}{3.67}\right) = 1360 \text{ Btu/mol}$$

$$\Delta S = \frac{1360}{492} = 2.76 \text{ Btu/mol-°R}$$

(b) Constant volume to final condition

$W = O$ and $Q = \Delta E$ (equation 9)

$\Delta E = \int C_v dt = 3/2\ R\ (984\text{-}492) = 1480$ Btu/mol

$\Delta H = \Delta E + \Delta PV = \Delta E + R\Delta T = 1480 + (1.99)\ (492)$
$= 2460$ Btu/mol

$$\Delta S = \int \frac{C_v}{T} dT = 3/2\ R \int \frac{dT}{T} = 3/2\ (1.99)\ (2.3) \left(\log_{10} \frac{984}{492}\right) = 2.08 \text{ Btu/mol-°R}$$

Path 3

(a) Adiabatic to final pressure
Q and $\Delta S = O$, therefore $\Delta H = -W$ for flow process
(If process had been a batch process, $\Delta E = -W$)
By equation 124

$$W = \frac{\gamma}{\gamma - 1} P_1 V_1 \left[\left(\frac{P_2}{P_1}\right)^{\frac{\gamma-1}{\gamma}} - 1\right]$$

The work done by the gas in expanding is

$$W = \frac{\gamma}{\gamma - 1} RT_1 \left[1 - \left(\frac{P_2}{P_1}\right)^{\frac{\gamma-1}{\gamma}}\right]$$

$\gamma = 5/3 = 1.67$; $\frac{\gamma}{\gamma-1} = \frac{1.67}{1.67-1} = 2.5$; $\frac{\gamma-1}{\gamma} = \frac{1}{2.5} = 0.4$

$W = (2.5)\ (1.99)\ (492)\ [1 - (.5)^{.4}] = 590$ Btu/mol

$\Delta H = -590$ Btu/mol

$\Delta E = \Delta H - \Delta PV = \Delta H - R\Delta T = -590 - 1.99\ (373\text{-}492) = -353$ Btu/mol

(b) Constant pressure to final volume.

$W = P\Delta V = R\Delta T = (1.99)\ (984\text{—}373) = 1215$ Btu/mol

$Q = \Delta H = \int C_p dT = 5/2\ R\ (984 - 373) = 3050$ Btu/mol

$\Delta E = \Delta H - \Delta PV = \Delta H - R\Delta T = 3050 - (1.99)\ (611)$
$= 1833$ Btu/mol

$$\Delta S = \int \frac{C_p}{T} dT = 5/2\ (1.99)\ (2.3) \left(\log_{10} \frac{984}{373}\right) = 4.84 \text{ Btu/mol °R.}$$

Path 4

(a) Adiabatic to final volume
Q and $\Delta S = O$

$$W = (2.5)\ (1.99)\ (492) \left[1 - \left(\frac{1.46}{14.7}\right)^{0.4}\right] = 1472 \text{ Btu/mol}$$

$\Delta H = -1472$ Btu/mol
$\Delta E = -1472 - 1.99\ (195 - 492) = -880$ Btu/mol

(b) Heat at constant volume to final temperature.

$\Delta E = 3/2R\ (984\text{—}195) = 2360$ Btu/mol

$\Delta H = \Delta E + R\Delta T = 2360 + (1.99)\ (984\text{—}195) = 3832$ Btu/mol

$$\Delta S = \int \frac{C_v}{T} dT = 3/2\ (1.99)\ (2.3) \left(\log_{10} \frac{984}{195}\right) = 4.84 \text{ Btu/mol °R}$$

The following tabulation summarizes these calculations:

TABLE 12

Solution to Alternate Paths for Perfect Gas Changes

PATH	W	Q	ΔE	ΔH	ΔS
		Btu/mol			Btu/mol °R.
1—a	676	676	0	0	1.38
b	980	2460	1480	2460	3.46
Σ	1656	3136	1480	2460	4.84
2—a	1360	1360	0	0	2.76
b	0	1480	1480	2460	2.08
Σ	1360	2840	1480	2460	4.84
3—a	590	0	—353	—590	0
b	1215	3050	1833	3050	4.84
Σ	1805	3050	1480	2460	4.84
4—a	1472	0	—880	—1472	0
b	0	3932	2360	3932	4.84
Σ	1472	3932	1480	2460	4.84

From the results given in Table 12 it can be seen that the changes in total energy, enthalpy and entropy are independent of the path taken. It can also be seen that the minimum work is for Path 2, where an iso-thermal change was followed as far as possible and the maximum work is for Path 1 where a constant pressure path was followed as far as possible.

Example 2

Problem: Derive a general expression for the change in entropy of 1 mol of perfect gas going from condition A to condition B.

Solution: Expressions will be derived for both batch and flow processes. The basic equation for both processes is

$$\Delta S = \int \frac{dQ}{T}$$

Batch Process

$$Q = \Delta E + W$$

$$\Delta E = \int_1^2 C_v dT$$

$$W = \int_1^2 P dV$$

$$Q = \int_1^2 C_v dT + \int_1^2 P dV$$

$$dQ = C_v dT + p dV = C_v dT + \frac{RT}{V} dV$$

$$\Delta S = \int \frac{dQ}{T} = \int_1^2 \frac{C_v dT}{T} + R \int_1^2 \frac{dV}{V}$$

$$\Delta S = C_v \ln \frac{T_2}{T_1} + R \ln \frac{V_2}{V_1}$$

Flow Process

$$Q = \Delta H + W$$

$$\Delta H = \int_1^2 C_p dT$$

$$-W = \int_1^2 V dP$$

$$dQ = C_p dT - \frac{RT}{P} dP$$

$$\Delta S = \int \frac{dQ}{T} = \int_1^2 \frac{C_p dT}{T} + R \int_2^1 \frac{dP}{P}$$

$$\Delta S = C_p \ln \frac{T_2}{T_1} + R \ln \frac{P_1}{P_2}$$

The analogy between the "Batch Process" and "Flow Process" entropy change equations is interesting and significant. Since $\frac{V_2}{V_1} = \frac{P_1}{P_2}$ for a perfect gas the only real difference between these equations for ΔS is the specific heats at constant volume, C_v, and constant pressure, C_p.

Example 3

Problem: A tank of 5 cubic feet capacity containing air at 1 atm. (abs.) and 25° C. is to be evacuated to 0.01 atm. (abs.) by means of a small reciprocating pump having a clearance of 2 percent and discharging to the atmosphere. On the assumption that the process as a whole is isothermal and reversible, what will be the change in entropy of the air during this process? What will be the change in entropy of the whole universe as a result of the occurrence of this process?

Solution: The 2 percent clearance has no bearing on the solution of this problem. Since the process is isothermal $\Delta E = O$ and $Q = W$, the required work being established by the specification that the process is reversible and isothermal. The net work is the algebraic sum of the work required to expand the air from the tank container into the cylinder of the vacuum pump and the work to expel the air from the cylinder with the piston. The latter work consists of two parts: (a) the work to compress the air in the vacuum pump cylinder from 0.01 to 1.0 atmosphere, and (b) the flow work to push the air from the cylinder.

Let N_1 = lb. mols of air in tank container at start.

N_2 = lb. mols of air remaining in tank container.

By perfect gas law

$$N_2 = \frac{(0.01)\ (14.7)\ (144)\ (5)}{(1543)\ (298)\ (1.8)} = 0.0001278$$

$N_1 = 100$ $N_2 = (100)\ (0.0001278) = 0.01278$

P_1 = initial pressure = 14.7 lbs./sq. in. abs.

P_2 = final pressure = 0.147 lbs./sq. in. abs.

In setting up the following work equations, a positive sign indicates work done by the air.

$$\text{Expansion work} = +N_1\ RT \ln \frac{P_1}{P_2} \quad \text{(equation 110)}$$

$$\text{Expulsion work} = -(N_1 - N_2)\ RT \ln \frac{P_1}{P_2} - (N_1 - N_2)\ RT$$

$$\text{Total work} = N_1\ RT \ln \frac{P_1}{P_2} - (N_1 - N_2)\ RT \ln \frac{P_1}{P_2} - (N_1 - N_2)\ RT$$

$$= N_2\ RT \ln \frac{P_1}{P_2} - (N_1 - N_2)\ RT$$

$$= (1.99)\ (298)\ (1.8)\ [(0.0001278)\ (2.3) \log_{10} 100 - 0.01265]$$

$Q = W = -12.9$ Btu

$$\Delta S = -\frac{12.9}{(298)\ (1.8)} = 0.024 \text{ Btu/°F.}$$

$\Delta S = O$ for universe since process is reversible.

Example 4

Problem: Usually the gas issuing from an oil well is saturated with water vapor. To reduce the pressure between the well-head and the gas pipe line, a throttle valve often is employed. This leads to difficulty due to ice forming in the valve with consequent diminution or stoppage of gas flow. To obviate this trouble, it is proposed to place several baffles in the line, thus increasing the friction and so maintaining the gas temperature above the freezing point of water. Explain the proposed change from a thermodynamic point of view, making recommendations.

Solution: In order to keep the water from freezing in the valve, sufficient heat must be added to maintain the temperature. The only way heat may be absorbed by the gas in such an installation is by radiation and convection from the atmosphere. When a gas is expanded it must either cool itself off or absorb heat from its surroundings. If the expansion takes place through a throttle valve there is not much time or space for the gas to absorb heat. If the expansion is allowed to take place more slowly by expanding through a series of baffles, a larger surface is available for heat absorption from the atmosphere. Consequently, the gas does not cool itself to as low a temperature as with a throttle valve. In the equation $Q - W = \Delta H$, $W = O$ for both types of chokes. Q is nearer zero for the throttle valve than for the baffled choke.

Example 5

Problem: A reciprocating air compressor having a clearance equal to two percent of displacement operates with adiabatic intake and exhaust. Compression and expansion may be assumed polytropic with $\gamma = 1.25$. If the machine operates without friction and requires five horsepower, when compressing from one atmosphere to four, what is its capacity in cubic feet per minute of free air at 70° F.?

Solution: The volumetric efficiency of this air compressor is found by means of equation 137 as follows:

$$E_v = \left[1.0 + 0.02 - 0.02 \left(\frac{4}{1}\right)^{\frac{1}{1.25}} \right] = 0.96$$

The volume of gas that a five-theoretical-horsepower compressor with this volumetric efficiency will handle is found by means of equation 133, as follows:

$$5 = \frac{144}{33{,}000}\ 14.7\ \frac{V_1}{0.96}\ \frac{1.25}{.25}\left[\left(\frac{4}{1}\right)^{\frac{.25}{1.25}} - 1\right]$$

$$V_1 = 46.7 \text{ cu. ft./min.}$$

The solution of a polytropic problem such as this one, which the value of γ is given, does not involve using the value of C_p/C_v or γ for isentropic compression. For air this adiabatic cofficient is 1.40. Had the mechanical efficiency of this machine been 85 percent the volume of air would have been about 40 cu. ft./min.

Example 6

Problem: Heat engines have been built using hot water as the working fluid. If hot water were available at 205° F. for use in such an engine, and cooling water at 65° F., what would be the minimum theoretical consumption of hot water per KWH generated by the hot water engine?

Solution: The maximum theoretical efficiency for any heat engine is the Carnot efficiency (see equation 46) which is dependent upon the highest and lowest temperatures only. For this particular case the efficiency of the Carnot cycle gives the ratio of work to heat added as

$$\frac{W}{Q} = \frac{205 - 65}{205 + 460} = 0.2105$$

Using the heat equivalent of one KW (3415 Btu) and assuming the specific heat of the hot water is 1.00 gives

$$\text{Lbs. of Hot } H_2O \text{ per KWH} = \frac{3415}{(0.2105)\ (205 - 65)} = 116$$

Example 7

Problem: Calculate the entropy change for the following processes:

(a) Evaporation of 1 lb. of water at 100 psia.
(b) Heating of 1 lb. of water at a constant pressure of 10 psi from 40° F. to 150° F.
(c) An adiabatic expansion of 1 lb. of steam from 600 lbs. psia and 700° F. to 25 psi and 240° F.
(d) A complicated series of processes using steam and starting at 1000 psi and 700° F., and ending at 400 psi and 560° F.

Solution: This problem is merely an exercise in reading the steam tables and Mollier diagram. (a) The entropy of vaporization for 1 lb. of H_2O at 100 psia is found to be 1.1287; (b) the entropies of water at 40° F. and 150° F. and 10 psi are essentially the same as for water at these temperatures under their vapor pressures of 0.122 and 3.716 psi because pressures have very little effect on the entropy or enthalpy of liquids, i.e., $\Delta S = 0.2149 - 0.0162 = 0.1987$ Btu/lb. ° F.; (c) this adiabatic expansion cannot be a reversible (constant entropy) one because the entropies at the initial and final conditions are different. The expansion may be considered as one that takes place in two steps; isentropic to final enthalpy, isenthalpic to final conditions, for which the entropy change is $\Delta S = S_2 - S_1 = 1.714 - 1.587 = 0.127$ Btu/lb. °F.; (d) entropy is a point function and changes in it are not dependent upon the path followed but only upon initial and final conditions, therefore $\Delta S = S_2 - S_1 = 1.5678 - 1.5138 = 0.0540$ Btu/lb. °F.

Example 8

Problem: Calculate the entropy change for the substance heated, for the heating material and for the system when 10 lbs. of water is heated in a coil heater from 60° F. to 150° F.

(a) By 2 lbs. of dry saturated steam at 1000 psia.
(b) By 2 lbs. of dry saturated steam at 50 psia.
(c) By 2 lbs. of superheated steam at 50 psia and 350° superheat temperature.
(d) In a perfect countercurrent heat exchanger using as a heat source 10 lbs. of water originally at 150° F.
(e) In a parallel flow heat exchanger using the minimum quantity of water originally at 200° F.

Solution: The entropy and enthalpy changes in 10 lbs. of water when it is heated from 60° F. to 150° F. are $\Delta S = 10\ (0.2149 - 0.0555) = 1.594$ Btu/°F. and $\Delta H = 10\ (117.87 - 28.07) = 898$ Btu. The entropy changes for the various cases are found as follows:

(a) the heat to be removed from each pound of the dry saturated steam at 1000 psi is $898 \div 2 = 449$ Btu which will leave 69.2% ($449 \div 649.5 = 0.692$) of the 1000 psi steam condensed for which $\Delta S = -(0.692)\ (0.6468) = -0.447$ Btu/lb. ° F. or -0.894 Btu/° F. The entropy change for the entire system is $\Delta S = 1.594 - 0.894 = 0.7$ Btu/° F.;

(b) fraction of 50 psi dry saturated steam that is condensed is $449 \div 923.9 = 0.486$, the entropy change being $\Delta S = -(0.486)\ (1.2473) = -0.606$ Btu/lb. °F. or -1.212 Btu/° F. $\Delta S_{total} = 1.594 - 1.212 = 0.382$ Btu/° F.;

(c) the following steam table readings are made for the solution of this part of the problem.

50 psia Steam	Saturated Water	Saturated Steam	Superheated Steam
t °F	281	281	350
H	250	1174	1210
S	0.411	1.658	1.705

This steam will lose its superheat and then partially condense and the fraction condensed is found by heat balance.

$$\text{Fraction condensed} = \frac{449 - (1210 - 1174)}{1174 - 250} = 0.447$$

$$\Delta S = -2\ [(1.705 - 1.658) + (0.447)\ (1.658 - 0.411)] = -1.208$$

$$\Delta S_{total} = 1.594 - 1.208 = 0.386 \text{ Btu/°F.};$$

(d) in a perfect countercurrent heat exchanger, the 10 lbs. of 150°F. water would be cooled to 60°F., for which $\Delta S_{total} = 1.594 - 1.594 = 0$, and the result would be a reversible process, which is obviously impossible; (e) in a parallel exchanger the heating media could not come out at a lower temperature than 150°. The amount of 200° water required would be

$$\frac{10\ (150 - 60)}{200 - 150} = 18 \text{ lbs.}$$

$$\Delta S = -18\ (0.2938 - 0.2149) = -1.42$$

$$\Delta S_{total} = 1.594 - 1.42 = 0.174$$

This simple problem illustrates approach to reversibility. The lower the temperature level of the source of the heat that is heating another body or doing work, the more nearly reversible the process is. The change in the entropy for the entire system is an index of this reversibility.

Example 9

Problem: (a) Tabulate the velocities which could be obtained by the isentropic expansion of steam from 300 psia and 200°F. superheat to 200 lb., 75 lb., 25 lb. and 3″ of Hg. pressure.

(b) Tabulate the velocities which could be obtained by adiabatic expansions to these same pressures, if the efficiencies of the nozzles were required for the expansion were 95 percent for the two highest back pressures, and 90 percent for the remainder.

Solution: In the solution of this problem use equation 29, which is

$$U_2 = 223.7 \sqrt{(1 - w)\ (H_1 - H_2)}$$

The following table gives the data read from the steam tables and also the calculated results.

P	t °F.	S	H	ΔH	Velocity, U_2 ft./sec. Without Friction	Value 1—W	With Friction
300	617	1.6345	1323.4	0	.....	.....	.....
200	530	1.6345	1280	43.4	1470	.95	1432
75	320	1.6345	1190	133.4	2580	.95	2515
25	240	1.6345	1105	218.4	3300	.90	3130
3″ Hg	115	1.6345	971	352.4	4200	90	3990

Example 10

Problem: It has been proposed to heat buildings in the following way:

"Refrigerating machines will be operated using the outside air as a heat source, and the warm air within the buildings as a heat sink. These refrigerating machines are to be operated by electrical power, thus pumping low-temperature heat uphill into buildings."

Making the assumptions listed below (based roughly on conditions in the San Francisco Bay region) calculate the maximum cost of electrical energy (cents per KWH) which would allow domestic heating by the proposed scheme to economically replace fuel heating.

(a) Refrigerating machines operate reversibly in every respect.

(b) Neglect all costs of refrigerating machines except the cost of power.

(c) Average winter temperature equals 40°F. An average room temperature equals 70°F.

(d) Average cost of fuel heat delivered to the rooms of a house is one dollar per million Btu.

Solution: The Carnot efficiency for the refrigeration (reversed heat engine) cycle is given by equation 47. For the temperatures involved in this problem the ratio of work required to the heat withdrawn from the cold reservoir, the atmosphere is

$$\frac{W}{Q_1} = \frac{70-40}{460+40} = 0.06$$

The heat delivered to the rooms $Q_2 = W + Q_1$

The work required per million Btu of heat delivered to the rooms is found as follows: $Q_1 = Q_2 - W = 1{,}000{,}000 - W$

$$\frac{W}{Q_1} = \frac{W}{1{,}000{,}000 - W} = 0.06$$

$W = 56{,}600$ Btu of work/million Btu of heat delivered

$$W = \frac{56{,}600}{3415} = 16.6 \text{ KW/million Btu of heat delivered}$$

Since fuel heat cost is $1.00 per million Btu, the minimum cost of electrical energy is

$$\frac{100}{16.6} = 6 \text{ cents per KWH}$$

Recently this method of heating has received considerable attention and is being tested in domestic installations.

An advantage of such a system is that the machine could be designed to run in the opposite direction in the summer time to cool the building.

Example 11

Problem: A boiler of 100 cubic feet capacity containing 3000 lbs. of water in the liquid phase, is being held at a constant pressure of 400 psia without withdrawing any steam from the boiler and 2000 lbs. of feed water at 150°F. are pumped in, while the pressure is meanwhile held constant at 400 lbs. How much heat is added to the boiler during this process?

Solution: The heat that is added to the boiler must be sufficient to maintain the temperature at 444.6°F. the temperature at which steam has a vapor pressure of 400 psi, while the additional boiler feed water is pumped in. Some of the steam in the vapor space must condense. The amount of steam in this space =

$$\frac{100 - (3000)\ (0.0193)}{1.1609} = 36.2 \text{ lbs. of sat. steam at 400 psi.}$$

Let X = the amount of lbs. of steam allowed to condense when the 2000 lbs. of feed water are added; then, neglecting changes of specific volume of water with temperature:

$$(3000 + 2000 + X)\ 0.0193 + (36.2 - X)\ 1.1609 = 100$$
$$X = 33.8$$

Heat to be added equals the heat required to heat the 2000 lbs. of boiler feed water to 444.6°F. minus heat liberated by condensing 33.8 lbs. of steam

$$Q = 2000\ (424 - 117.9) - (33.8)\ (780.9) = 585{,}800 \text{ Btu.}$$

Example 12.

Problem: A direct-fired superheater is fed simultaneously with steam from two lines, the first carrying dry saturated steam at 100 psia, and the second carrying steam at 120 psia and 90 percent quality. At the coil exit the pressure of the combined stream is 90 psia and the steam has 100°F. superheat. During passage through the coil 120 Btu per pound of total steam flowing were added. What weight percent of the total steam fed comes from the 100-lb. line?

Solution: This is an exercise in using the steam tables and making a heat balance. Based on a total of 100 pounds of steam:

Let x = weight % of steam from the 100-lb. line; then $100 - x$ = weight % of steam from 120-lb. line. The following heat balance is written on the superheater to solve for x.

$$x\ (1187.3) + (100 - x)\ (312.46 + 0.900 \times 878.1) + (100)\ (120) = (100)\ (1239.1)$$
$$x = 20\%$$

Example 13.

What is the entropy change for the mixing of 100 pounds of lead shot originally at 250°F. with 10 pounds of water originally at 60°F. The specific heat of lead is 0.03.

Solution: The final temperature after mixing is first found by a heat balance as follows:

$$(100)\ (0.03)\ (250 - t) = (10)\ (1.0)\ (t - 60)$$
$$t = 104°\text{F.}$$

$$\Delta S_{\text{lead}} = \int \frac{dq}{T} = (100)\ (0.03)\ (2.3)\left(\log_{10}\frac{460+104}{460+250}\right)$$

$$\Delta S_{\text{lead}} = -0.69 \text{ Btu/°F.}$$

$$\Delta S_{\text{water}} = (10)\ (1.0)\ (2.3)\left[\log_{10}\left(\frac{460+104}{460+60}\right)\right]$$

$$\Delta S_{\text{water}} = 0.82$$

$$\Delta S_{\text{system}} = 0.82 - 0.69 = 0.13 \text{ Btu/°F.}$$

Example 14.

Problem: For an ammonia compression refrigeration system operating between an 86°F. condenser and a 5°F. evaporator, compute the following; (a) coefficient of performance (ratio of heat removed to work supplied); (b) horsepower per ton (200 Btu/min. heat removal); and (c) ammonia circulation per ton.

Solution: The following data are read from the Bureau of Standards Circular No. 142 on Ammonia.

			Enthalpy Btu/lb.	
	T°F.	P psia	Sat. liquid	Sat. vapor
Condenser	86	169.2	138.9	—
Evaporator	5	34.27	—	613.5

The compression follows a constant entropy path from sat. vapor @ 5°F. and 34.27 psia to 169.2 psia giving an enthalpy of 713 Btu/lb. at the end of the compression path.

Refrigeration, Btu/lbs. $NH_3 = 6135 - 138.9 = 474.4$

Work of Compression, Btu/lb. $NH_3 = 713 - 6135 = 99.7$

Cooling Requirements for condensing Btu/lb. $NH_3 = 713 - 138.9 = 574.1$

$$\text{Coefficient of Performance} = \frac{474.4}{99.7} = 4.76$$

Ammonia circulation per minute for one ton of Refrigeration $= \frac{200}{474.4} = 0.42$ lbs. NH_3

$$\text{Horsepower/ton of refrigeration} = \left(\frac{12000}{2546}\right)\left(\frac{99.7}{474.4}\right) = 0.985$$

Example 15.

Problem: A combined expander-compressor machine operating on the same shaft may be used to sub-cool a fluid below the temperature conveniently reached by available cooling facilities, such as cooling water, refrigerants, etc. In this type of machine the heat is removed from the fluid in the form of work, which may be used to compress gas or it may be converted into power. Machines of this kind have been used in liquefying air. Before being expanded the fluid must be compressed and cooled to a point above the two-phase border curve so part of it will condense upon expansion. If this expansion is an ideal frictionless process, the path followed will be isentropic. In actual practice there is some increase in entropy, resulting in a higher temperature than could be reached by an isentropic expansion. This method of subcooling a fluid might be used advantageously in liquefying methane, for example. As an illustration of this process, consider CO_2 at 1000 psia and 100°F. (conditions that could be reached by multistage compression and cooling). Find lowest possible temperature fluid could be cooled to by isentropic expansion to 200 lbs. and the amount liquefied.

Solution: Isentropic expansion of CO_2 from an initial point of 1000 psia and 100°F. to a pressure of 200 psia involves removal of $278 - 255 = 23$ Btu/lb. of work, reaching a temperature of —25°F. and liquefaction of 20% of the original fluid. These figures were obtained from the CO_2 Mollier diagram in Dodge's "Chemical Engineering Thermodynamics."

Some of this work may be recovered as power and used to recompress the uncondensed gas part way back to the initial conditions.

Non-Isothermal Flow of Compressible Fluids

The usual fluid flow calculations are for conditions that may be assumed isothermal, at an average temperature, and for fluids that may be assumed noncompressible, at an average density. Problems of this kind involve application of the Fanning equation and usually require no kinetic energy computations. The flow of compressible fluids (perfect and actual gases) through heated conduits is accompanied by changes in temperature and density of such magnitude that it is necessary to make refinements in the methods of computation.

There are two methods that may be employed. In both methods, the gas density is replaced by the gas PVT equation, and the temperature is replaced by a heat balance equation. The difference between the methods is the way the kinetic energy term is handled. In the simpler method the kinetic energy difference is neglected while calculating wall friction and then included later. In the more rigorous method

one comprehensive equation is developed to include both friction and kinetic energy. The development of these equations follow.

The starting point in deriving both methods is Bernoulli's theorem, which is given by equation 33, repeated here for convenience:

$$-\int VdP + W = \Delta X + \frac{\Delta U^2}{2g} + F$$

For the flow conditions being considered the W and ΔX terms equal zero giving

$$-\int VdP = \frac{\Delta U^2}{2g} + F \qquad (139)$$

Differentiating gives

$$-VdP = \frac{UdU}{g} + dF \qquad (140)$$

Equation 140 is general and is the basis for the development of the two calculation methods described above.

Another relationship required is the gas PVT equation

$$PV = Z\,RT \qquad (141)$$

Differentiating equation 141

$$dV = \frac{ZR}{P}\,dT - \frac{ZRT}{P^2}\,dP \qquad (141A)$$

where P = pressure, lbs./sq. ft.
V = specific volume, cu. ft./lb.
T = temperature ° R
R = gas constant, 1543/mol wt.

In equation 141, Z is the deviation from the ideal gas law (PV = RT) and, as such, varies with T and P. For these purposes an average Z value may be used. Carrying Z along as a dependent variable complicates the equation unnecessarily.

The third relationship required in formulating the non-isothermal flow equations for compressible fluids is a heat balance

$$WC_p dT = Q\pi D\,dL \qquad (142)$$

$$dT = \frac{Q\pi D}{WC_p}\,dL \qquad (142A)$$

Upon integrating

$$T = T_o + \frac{Q\pi D}{WC_p}\,L \qquad (143)$$

where W = weight rate of flow, lbs./hr.
C_p = specific heat at constant pressure, Btu/lb.—°F.
Q = rate of heat input to gas, Btu/hr.—sq. ft. of heated surface.
D = diameter of conduit, feet
T = temperature °R
T_o = initial temperature °R
L = conduit length, feet

Equation 142 is valid for heat addition at a uniform rate per unit heated area.

The last fundamental relationship required in this equation development is the Fanning equation for the wall friction

$$F = \frac{2fU^2L}{gD} \qquad (144)$$

but $U = \frac{WV}{A}$, which upon substitution in equation 144 gives

$$F = \frac{2f\,W^2V^2L}{g\,D\,A^2} \qquad (145)$$

where
F = friction head in feet of fluid flowing
f = friction factor and function of Reynolds number
U = velocity in feet/sec.
D = diameter of conduit in feet
A = area of conduit in square feet.
g = gravitational constant

From the above equations, expressions for non-isothermal flow of a compressible fluid will next be developed.

Case I Neglecting Kinetic Energy Change

If the kinetic energy term is neglected, equation 140 becomes

$$-VdP = dF \qquad (146)$$

Substituting for F from equation 144,

$$-dP = \frac{2f\,W^2V}{g\,DA^2}dL$$

Eliminating V by equation 141

$$-PdP = \frac{2f\,W^2Z\,RT}{g\,DA^2}\,dL$$

Eliminating T by equation 143

$$-PdP = \frac{2f\,WZR}{gA^2}\left[\frac{WT_o}{D} + \frac{Q\pi L}{C_p}\right]dL \qquad (147)$$

Integrating equation 147 gives the following expression for the wall friction

$$P_o^2 - P^2 = \frac{4fWZRL}{g\,A^2}\left[\frac{WT_o}{D} + \frac{Q\pi L}{2C_p}\right] \qquad (148)$$

Where:

T_o and P_o are initial conditions at start of conduit, where L = o Equation 148 gives wall friction ΔP. To this must be added the velocity head change $\Delta\frac{U^2}{2g}$.

Case II Including Kinetic Energy Term

This development starts with equation 140. The dF term is handled as above. The UdU term is set up as follows

$$U = \frac{WV}{A}$$

$$dU = \frac{W}{A}dV$$

$$UdU = \frac{W^2}{A^2}VdV \qquad (149)$$

Combining equations 140, 145 and 149

$$-VdP = \frac{W^2VdV}{A^2g} + \frac{2fW^2V^2}{g\,DA^2}\,dL$$

Dividing through by $\frac{VW^2}{gA^2}$ and combining with equation 141 and 141A

$$-\frac{gA^2}{ZRW^2}\,PdP = dT - \frac{T}{P}\,dP + \frac{2f\,T}{D}\,dL \qquad (150)$$

Eliminating T and dT by means of equations 142A and 143

$$-\frac{gA^2}{ZRW^2}\,PdP = \frac{Q\pi D}{WC_p}\,dL - \left(T_o + \frac{Q\pi DL}{WC_p}\right)\frac{dP}{P} + \frac{2f}{D}\left(T_o + \frac{Q\pi DL}{WC_p}\right)dL \qquad (151)$$

Equation 151 may be put in the following form

$$-\frac{dP}{dL} = \frac{1 + \frac{2f}{D}\left(\frac{T_oWC_p}{Q\pi D} + L\right)}{\frac{gA^2C_pP}{ZRWQ\pi D} - \left(\frac{\frac{T_oWC_p}{Q\pi D} + L}{P}\right)} \qquad (152)$$

This equation is too complex to obtain a straight forward integrated expression. Accordingly, it must be solved stepwise, computing the pressure drops for each successive incremental length from the flowing conditions and the cumulative pressure drop up to the point in question.

The "f" and "Z" terms in equation 152 vary with conditions. In some cases these variations might be of such magnitude to require evaluating "f" and "Z" at each increment, which increases the calculation time somewhat but this is better than trying to formulate these as functions of flowing conditions. The friction factor could be expressed as a simple function of Reynolds' number but the gas compressibility factor is much more complex.

Example 16

Problem: In order to illustrate and compare the above two methods, the pressure drop for the following problem will be computed both ways:

Fluid: 2000 lbs./hour of air
Conduit: 12 feet of 2.5-inch schedule 40 steel pipe
Inlet Conditions: Temperature of 100° F. and pressure of 50" H_2O above atmospheric pressure of 14.7 psia.
Heat Flux: 45,000 Btu/hr. — ft.² of heated surface.

Find the temperature and pressure at the outlet of the conduit. For this problem the fluid behaves as a perfect gas, i.e., PV = RT.

Solution by Approximate Method: From pipe tables the dimensions of the conduit are found to be

Internal diameter, D = 2.469" or 0.2056'
Flow Area, A = 0.03322 sq. ft.
Heated Perimeter, πD = 0.647'

The specific heat of air at constant pressure is 0.241 Btu/lb. — °F. The friction factor will be evaluated at average conditions.

Initial Conditions are found as follows:

$$\text{Inlet pressure} = (14.7)\ (144) + (50)\left(\frac{62.4}{12}\right) = 2116 + 260 = 2376 \text{ lbs./sq. ft.}$$

$$\text{Air density} = \left(\frac{29}{359}\right)\left(\frac{492}{560}\right)\left(\frac{2376}{2116}\right) = 0.0797 \text{ lbs./cu. ft.}$$

$$\text{Velocity, } U = \frac{2000}{3600\ (0.0797)\ (0.03322)} = 210 \text{ ft./sec.}$$

$$\text{Kinetic energy} = \frac{(210)^2}{64.4} = 685 \text{ ft. lbs./lb. air}$$

$$\text{Inlet velocity head} = (685)\ (0.0797) = 54.6 \text{ lbs./ft.}^2$$

$$54.6\left(\frac{12}{62.4}\right) = 10.5 \text{ inches/H}_2\text{O}$$

The initial stream pressure available to push the gas through the pipe is obtained by subtracting this inlet velocity head from the static pressure

$$P_o = 2376 - 54.6 = 2321.4 \text{ lbs./ft.}^2$$

Final conditions are found as follows by assuming a pressure drop. If the calculated pressure drop is different, this calculation must be repeated with a revised pressure.

Assume a stream pressure of one atmosphere at end of pipe

Final stream pressure = 2116 lbs./sq. ft.

Final air temperature (from equation 143)

$$t = 100 + \frac{45{,}000\ (0.647)\ (12)}{2000\ (0.241)} = 825°\text{ F.}$$

$$T = 825 + 460 = 1285°\text{R}$$

$$\text{Air density} = \left(\frac{29}{359}\right)\left(\frac{492}{1285}\right)\left(\frac{2116}{2116}\right) = 0.0309 \text{ lbs./cu. ft.}$$

$$\text{Velocity} = \frac{2000}{3600\ (0.0309)\ (0.03322)} = 542 \text{ ft./sec.}$$

$$\text{Kinetic Energy} = \frac{(542)^2}{64.4} = 4560 \text{ ft. lbs./lb. gas}$$

$$\text{Final Velocity Head} = (4560)\ (0.0309) = 141 \text{ lbs./ft.}^2$$

$$141\left(\frac{12}{62.4}\right) = 27.1 \text{ inches H}_2\text{O}$$

This final velocity head is available for recovery. The percentage of it recovered will depend upon the shape of the pipe exit, etc.

The acceleration loss may be estimated by multiplying the increase in the kinetic energy term by the average density of the fluid. The logarithmic average density will be used.

$$\rho_1 = 0.0797$$
$$\rho_2 = 0.0309$$
$$\rho_{av} = 0.051$$

$$\text{Acceleration Loss} = 0.051\ (4560 - 685) = 198 \text{ lbs./ft.}^2$$

$$\text{Acceleration Loss} = 198\ \frac{12}{62.4} = 38 \text{ inches H}_2\text{O}$$

The wall friction is found by equation 148. The friction factor, f, is found to be 0.0037

$$P_o^2 - P^2 = \frac{4(0.0037)\ (2000)\ (1.0)\left(\frac{1543}{29}\right)(12)}{(32.2)\ (3600)^2\ (0.03322)^2}$$

$$\left[\frac{(2000)\ (560)}{0.2056} + \frac{(45000)\ (3.1416)\ (12)}{2\ (0.241)}\right] = 369{,}000$$

$$P = \sqrt{(2321.4)^2 - 369{,}000}$$

$$P = 2240 \text{ lbs./ft.}^2$$

$$\Delta P = \frac{2321.4 - 2240}{5.2} = 15.5 \text{ inches H}_2\text{O}$$

Total pressure drop = 15.5 + 38 = 53.5 inches H_2O, which checks the assumed pressure used in the calculations.

Solution by Differential Equation: Equation 152 is applied to the same problem by one foot increments. The initial and final conditions computed in the first part are used in setting up the following equation

$$-\frac{dP}{dL} = \frac{\left[1 + \left(\frac{2(0.0037)}{0.2056}\right)\left(\frac{(560)\ (2000)\ (0.241)}{(45{,}000)\ (0.647)} + L\right)\right]}{\left[\frac{(32.2)\ (3600)^2\ (0.03322)^2\ (0.241)P}{(1.0)\ \frac{1543}{29}\ (2000)\ (45{,}000)\ (0.647)}\right] - \left(\frac{(560)\ (2000)\ (0.241)}{(45{,}000)\ (0.647)} + L\right)}$$

$$-\frac{dP}{dL} = \frac{1.334 + 0.036L}{0.000036\ P - \left(\frac{9.27 + L}{P}\right)}$$

This equation will be evaluated by one foot increments, finding dP/dL from the above equation, calling this $\Delta P/\Delta L$, subtracting the ΔP increments from pressure to get new P values for the next step.

Feet	0.036L	1.334+ 0.036L	9.27+L	P lbs/ft²	9.27+L / P	0.000036P	ΔP/ΔL lbs./sq. ft/ft
0	0	1.334	9.27	2321.40	0.00399	0.0836	−16.75
1	0.036	1.370	10.27	2314.55	0.00444	0.0830	−17.44
2	0.072	1.406	11.27	2287.21	0.00492	0.0823	−18.17
3	0.108	1.442	12.27	2263.04	0.00540	0.0817	−18.90
4	0.144	1.478	13.27	2250.14	0.00589	0.0810	−19.66
5	0.180	1.514	14.27	2230.48	0.00640	0.0803	−20.25
6	0.216	1.550	15.27	2210.23	0.00691	0.0795	−21.35
7	0.252	1.586	16.27	2188.88	0.00742	0.0789	−22.48
8	0.288	1.622	17.27	2166.40	0.00797	0.0780	−23.20
9	0.324	1.658	18.27	2143.20	0.00851	0.0773	−24.05
10	0.360	1.694	19.27	2119.15	0.00911	0.0761	−25.25
11	0.396	1.730	20.27	2093.90	0.0097	0.0753	−26.35
Outlet				2067.55			

Total ΔP = 2321.4 − 2067.55 = 253.85

$\Delta P = 253.85\left(\frac{12}{62.4}\right) = 48.8''$ H_2O

Conclusions: The above two solutions give a total pressure drop of 53.7 inches of H_2O by the approximate method and 48.8 inches of H_2O by the stepwise application of the differential equation. The approximate method involved using an average density, whereas the differential equation method did not. A logarithmic average density was used in preference to an arithmic average because the logarithmic density gave better agreement with the more precise stepwise method. The difference could be more significant.

This same technique has applications in hydrocarbon processing, such as heating a gas in a tubular fired heater.

(End of Part VI; Part VII will appear in the next issue.)

The Aromatic Adsorption Index as a Rapid Method for Approximating Catalyst Activity

W. W. SCHEUMANN and A. R. RESCORLA
Cities Service Refining Corporation, Lake Charles, La.

CONTROL of catalytic cracking operations requires frequent measurement of catalyst activity. Usually such measurements are made by running a standardized laboratory catalytic cracking test, so controlled that the catalyst activity is the only variable. This requires considerable equipment, and is time-consuming.

The "aromatic adsorption index" is based on the ability of a cracking catalyst selectively to adsorb aromatic hydrocarbons from a hydrocarbon mixture. This test can be made in a fairly short time, and does not require equipment other than the usual equipment of most laboratories. In this paper the authors relate the aromatic adsorption index with the D + L† activity for natural and synthetic catalysts.

This paper was presented in November to a Division of Refining Symposium on Cracking Catalysts during the annual API meeting in Chicago.

THE activity of the catalyst employed in the catalytic cracking of petroleum fractions is of major importance in that, for a given feed stock, it dictates the severity of cracking conditions necessary to achieve the desired conversion. Catalyst-activity measurements are, therefore, a necessary part of the laboratory work required for satisfactory control of such cracking operations.

Various test methods for evaluating the activity of a catalyst have been proposed and used by the industry. These procedures fall into two principal groups: One group of methods is based on a standardized laboratory cracking operation, and the most widely used procedures of this type include the UOP activity test,[1] the Houdry catalytic activity index,[2] the Jersey D + L activity test,[3] and the Kellogg fluidized fixed-bed activity test.[4] In contrast with these methods based on a laboratory cracking procedure, the second type of test method for catalyst-activity measurement is based on surface-area reactions. Three common methods of this type are the potassium hydroxide (KOH) adsorption test,[5] the aromatic adsorption index,[6] and the nitrogen adsorption surface-area measurement.[7]

In the authors' laboratories the standard procedure for activity determinations for catalysts used in fluid catalytic cracking operations has been the Kellogg fluidized fixed-bed activity test. However, for a rapid test method or for a check test of results by the standard method the aromatic adsorption index has also been used. This paper presents data showing the degree of correlation between the two methods, on both natural and synthetic fluid catalysts, based on samples taken during actual commercial operations.

† Distillate plus loss.

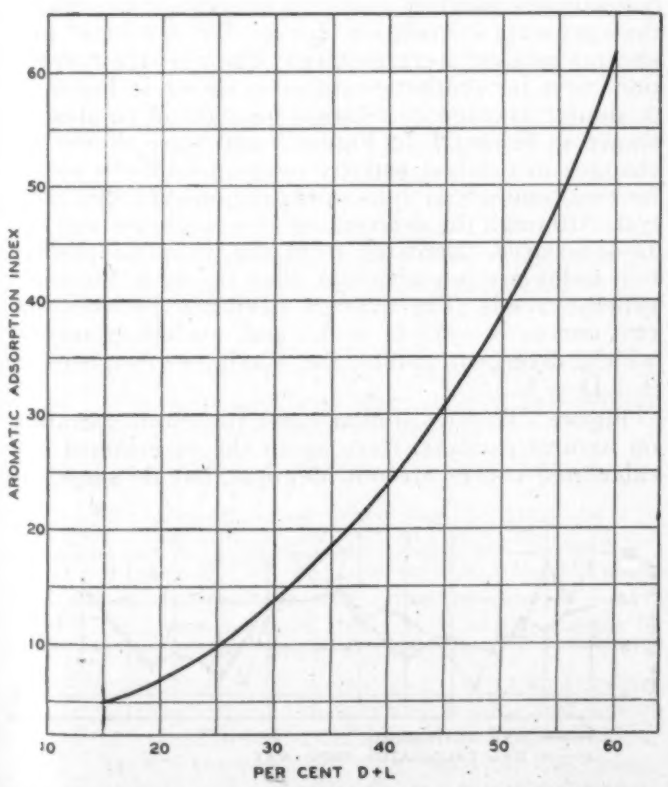

Figure 1. Relation of Aromatic Adsorption Index and Percent D + L for Synthetic Catalyst.

Test Procedures

The Kellogg fluidized fixed-bed activity test is carried out by actually cracking a standard East Texas

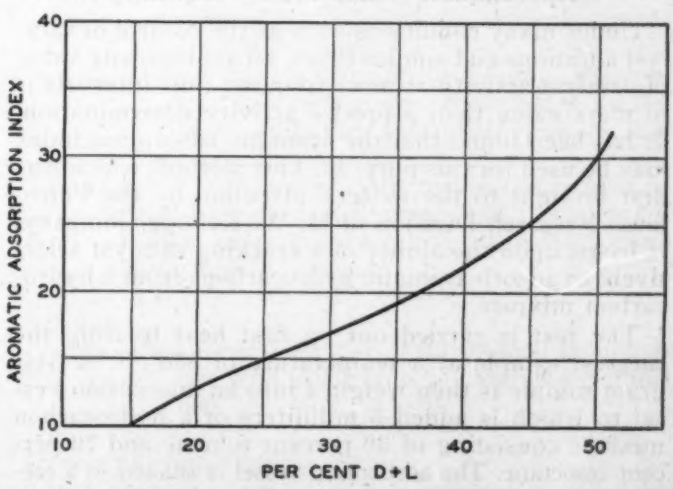

Figure 2. Relation of Aromatic Adsorption Index and Percent D + L for Natural Catalyst.

gasoil under a set of standardized and controlled operating conditions in a specifically designed small laboratory cracking unit, in which the test catalyst is maintained as a fluidized bed. The stabilized liquid products are tested by ASTM method D86-46, and the percent evaporated at 400° F. is expressed as percent D + L (distillate plus loss). By carrying out all tests under standard conditions of operation and using the same feed stock, the activity of the catalyst remains as the only variable, and the production of distillate plus loss is a measure of the activity of the catalyst.

Prior to the actual cracking operation, the catalyst must be elutriated to remove a majority of particles of less than 20 microns average size, as such sized particles do not aerate satisfactorily in the test unit. New catalyst samples require a heat treatment before testing, and used catalyst must be burned to a carbon content of less than 0.1 percent for consistent results.

In addition to the activity expressed as per cent D + L, the measurement of the non-condensible gas volume permits calculation of the gas-producing factor for the catalyst and a determination of the carbon laid down on the catalyst during the test, a calculation of the carbon-producing factor.

This method of testing yields, therefore, much valuable information regarding the characteristics of a catalyst. It is, however, time-consuming, as an average of 10 to 12 man-hours of work and a total elapsed time of 20 to 24 hours is required for each sample. Moreover, a very considerable amount of specialized equipment is necessary if multiple units of some type are set up in order to decrease the man-hour requirements per determination.

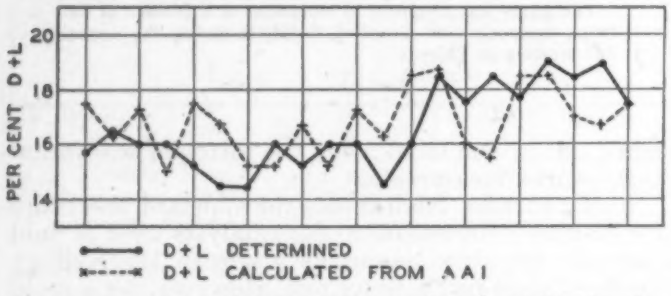

Figure 3. Comparison of Activity Values on Commercial-Unit Samples over Eight-Month Period—Synthetic Catalyst.

Approximate Value More Frequently

Under many conditions such as the control of catalyst additions and similar items, an approximate value of catalyst activity at more frequent time intervals is of more value than a precise activity determination. It has been found that the aromatic adsorption index may be used for this purpose. This method, which was first brought to the writers' attention by the Petroleum Research Division of M. W. Kellogg Company, is based upon the ability of a cracking catalyst selectively to adsorb aromatic hydrocarbons from a hydrocarbon mixture.

The test is carried out by first heat treating the catalyst sample at a temperature of 850° F. A 3½-gram sample is then weighed into an adsorption vessel to which is added 5 milliliters of a hydrocarbon mixture consisting of 30 percent toluene and 70 percent *iso*octane. The adsorption vessel is shaken in a mechanical shaker for two hours, after which the refractive index is determined on the supernatant liquid.

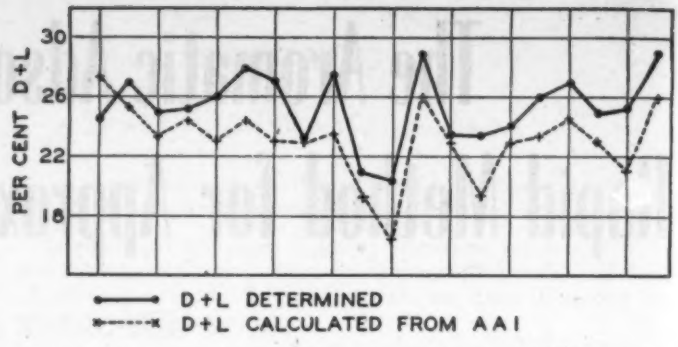

Figure 4. Comparison of Activity Values on Commercial-Unit Samples over Eight-Month Period—Synthetic Catalyst.

The aromatic adsorption index is the difference between the refractive index of the original hydrocarbon mixture and that of the hydrocarbon mixture after contact with the catalyst multiplied by 10.[4] A detailed description of the Cities Service modification of the original Kellogg procedure is given in the appendix.

This determination requires approximately five hours' elapsed time, and it is quite apparent that a single laboratory operator can carry out a considerable number of such determinations simultaneously, so that the man-hour requirements per determination is quite small. In addition, the only piece of expensive laboratory equipment required is a precision refractometer equipped for constant-temperature work.

Relation of D + L to Aromatic Adsorption Index

In order to relate the aromatic adsorption index to the D + L value, a very considerable number of samples of new and regenerated synthetic catalyst and new and regenerated natural catalyst from both commercial units and pilot-plant operations were tested by the two methods. This program was carried out over an extended time period, and included catalyst samples varying widely in activity. It was found that separate correlation curves for synthetic and natural catalyst were required. The average correlation curve for synthetic catalyst is shown in Figure 1. A similar average correlation for natural catalyst is shown in Figure 2. In Figure 3 and 4 are shown the changes in catalyst activity over an 8-month period for two commercial units operating on synthetic catalyst. Although the determined D + L curves and the D + L curve calculated from the aromatic adsorption index are not identical, they do show the same general trends. The average deviation between the two curves is —0.8 D + L; and, excluding several widely divergent points, the maximum deviation is ± 2 D + L.

Figure 5 shows a similar curve for a unit operating on natural catalyst. Here again the determined and calculated curves are not identical, but do show the

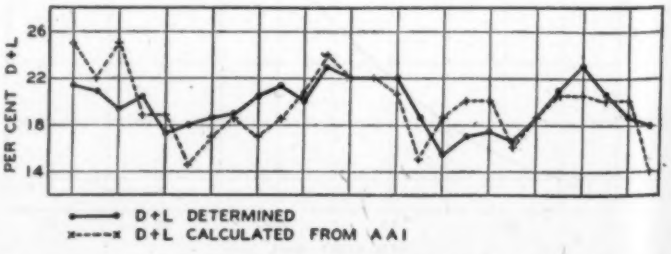

Figure 5. Comparison of Activity Values on Commercial-Unit Samples over Eight-Month Period—Natural Catalyst.

same general trends. The average deviation is —0.3 D + L; and, excluding several widely divergent points, the maximum deviation is ± 2 D + L.

Conclusions

Although the aromatic adsorption index does not in all cases measure catalyst activity precisely in agreement with the fluidized fixed-bed D + L method, it does give a fairly good approximation. It should have many applications where an approximate measurement of activity is suitable, and where results are needed in a short elapsed time interval, or where relative changes in activities of a large number of samples is desired.

Acknowledgment

The authors are appreciative of the assistance obtained from the Cities Service Refining Corporation laboratory staff in obtaining the data used in this article.

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APPENDIX

Laboratory Procedure for Determination of Aromatic Adsorption Index

(Cities Service Refining Corporation Modification of The M. W. Kellogg Company Method)

Purpose

1) Because silica-gel type catalysts will selectively adsorb aromatics from mixtures of hydrocarbons, an attempt is being made to estimate the activity of catalysts by measuring the extent to which this adsorption will take place under controlled conditions.

2) The aromatics adsorption index has an advantage over D + L or similar activity ratings in that the personnel requirements are much smaller, and the test may be completed in a shorter period of time.

Equipment

3) (a) Bausch and Lomb precision refractometer equipped with constant-temperature apparatus.
   (b) Apparatus: 5-millititer pipettes and micro pipettes.
   (c) Adsorption bottles: Use glass-stoppered weighing bottles 4 centimeters high and 2.5 centimeters in diameter.
   (d) Shaker capable of shaking the samples continuously for two hours.
   (e) Muffle furnace equipped with means of measuring and controlling the temperature.

Reagents

4) Toluene-*iso*octane solution: 30 percent by volume toluene in *iso*octane. Use reagent-grade toluene and reference-fuel grade *iso*octane.

Procedure

5) If the catalyst is new, place about 10 grams of it in a crucible and heat it in the muffle furnace for 2 hours at 850° F. If the catalyst has been used, heat the crucible containing the catalyst for at least two hours at 1000° F. Cool the sample to room temperature in a desiccator.

6) Weigh duplicate 3.50-grams ± 0.1-gram portions of the catalyst into weighing bottles and pipette in 5.0 milliliters of the toluene-*iso*octane solution. Tightly stopper each bottle, place it in the shaker, and shake for two hours at room temperature. Investigation has shown that, in most cases, equilibrium will be approached in that length of time.

7) Allow the catalyst to settle to the bottom of the bottle; and, using a micro pipette, remove a portion of the clear liquid and determine its refractive index. It is well to repeat the refractive index two or three times on separate portions of the liquid, if possible.

8) Determine refractive index of the original toluene-*iso*octane solution.

9) Occasionally a catalyst is found which absorbs so much of the toluene-*iso*octane solution that there is insufficient clear liquid available for a refractive-index determination. In such cases 2.50 grams ± 0.01 gram of catalyst with the 5 milliliters toluene-*iso*octane solution should be used, and the special formula employed for calculating the aromatics adsorption index.

10) All refractive indices are measured at 20° C. using the sodium "D" line.

Calculations

11) The aromatics adsorption index, *I*, is defined as follows:

$$I = (n^{\circ}_D - n'_D)\ (10^4)$$

Where:

n°_D = the refractive index of the original toluene-*iso*octane solution.

n'_D = the refractive index of the solution after contact with 0.70 grams of catalyst per milliliter until the adsorption process has reached an equilibrium.

12) If it has been necessary to reduce the amount of catalyst sample to 2.50 grams, the aromatics adsorption index may be calculated by the following formula:

$$I = 1.41\ (n^{\circ}_D - n''_D)\ (10^4) - 1.1$$

Where:

n''_D = the refractive index of the solution after contact with 0.50 grams of catalyst per milliliter.

Accuracy of Reporting

13) Report the aromatics adsorption index to the nearest 0.1 unit.

High Range Vapor Pressures and Antoine Equation Constants of Some C_3 to C_5 Hydrocarbons

JOHN GRISWOLD and W. B. BROOKS
The University of Texas, Austin

LITERATURE values for high range vapor pressures of seven of the more common C_4 hydrocarbons found in commercial materials and of propene, propane, iso- and n-pentane, are correlated and presented in tabular form.

A straightforward procedure for determining Antoine equation constants based on the shape of typical log P versus 1/T curves gives an improved fit to the vapor pressure data of these hydrocarbons.

Fractionation of light hydrocarbons under pressure is essential to many processes in the manufacture of natural, motor, and aviation gasoline, of synthetic rubber, and of chemical products.

For mixtures of close-boiling hydrocarbons, tower plate calculations require vapor pressure data of a high degree of accuracy. Although absolute accuracy is most desirable, it is actually subordinate to *interconsistency* of vapor pressures between different compounds. The present work extends the vapor pressure correlation of C_4 hydrocarbons recently presented by Hachmuth, Hanson and Smith.[5]

NUMEROUS mathematical relations have been used to represent vapor pressure data. The various equations differ greatly in their complexity and in the accuracy and range over which they fit experimental data. In general, the more complex the equation, the more accurately it can be made to fit wide-range data; but an accurate, simple equation is highly

TABLE 1

Boiling Points, Selected Criticals, and Constants for Zero-Residual Vapor Pressure Equation (log P = B — A/T°R.)

	T_b °F.[a]	T_c °F.[b]	P_c psia[b,c]	A	B
Propene	—53.86	197.4	667.4	1760.858	5.502902
Propane	—43.73	206.3	618.9	1801.854	5.495858
Isobutane	10.89	274.9	529.2	2039.956	5.499490
Iosbutene	19.58	292.5	580.7	2111.174	5.569453
1-Butene	20.73	297.0	588.0	2110.306	5.557091
1, 3-Butadiene	24.06	305.6	627.7	2146.147	5.600956
n-Butane	31.10	305.6	551.3	2155.941	5.557352
Trans-2-Butene	33.58	311.0	595.4	2205.074	5.634797
Cis-2-Butene	38.70	320.0	610.1	2237.407	5.653833
Isopentane	82.14	370.0	482.2	2369.538	5.538049
n-Pentane	96.93	386.8	485.1	2470.614	5.603428

[a] Converted from "Selected Values of Properties of Hydrocarbons."
[b] From LC736, Nat'l Bur. Stds., and other sources noted in context.
[c] Conversions from reference values listed to 0.1 atm. tabulated to 0.1 psia for consistency with Table 2.

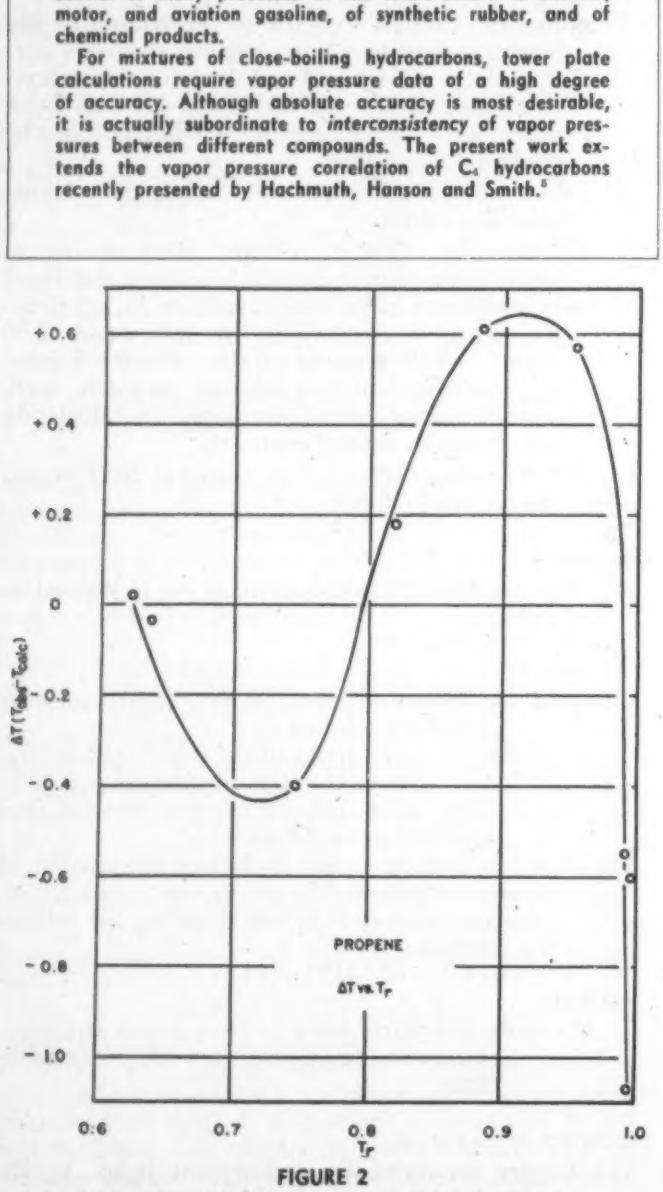

FIGURE 2
Temperature residuals for propene, from Equation 2.

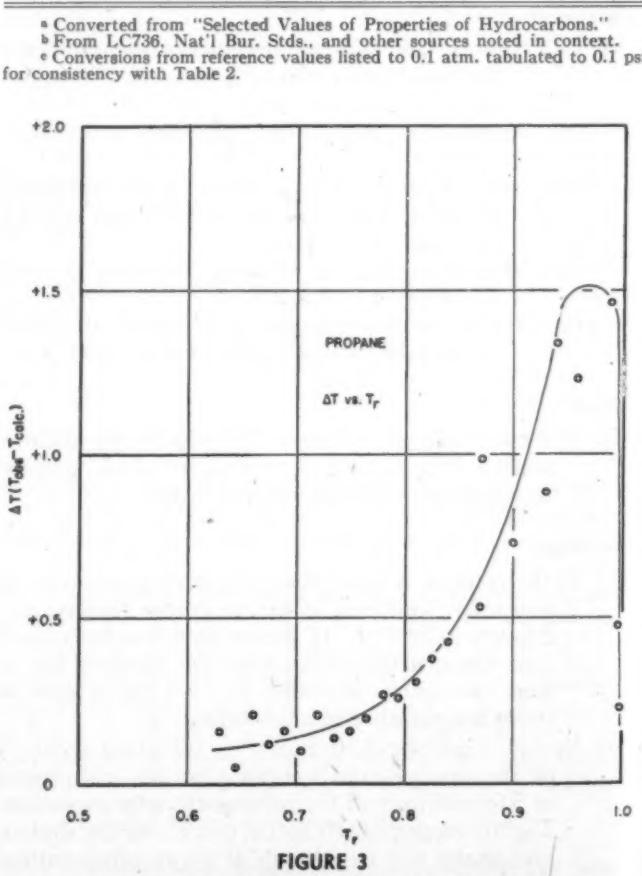

FIGURE 3
Temperature residuals for propene, from Equation 2.

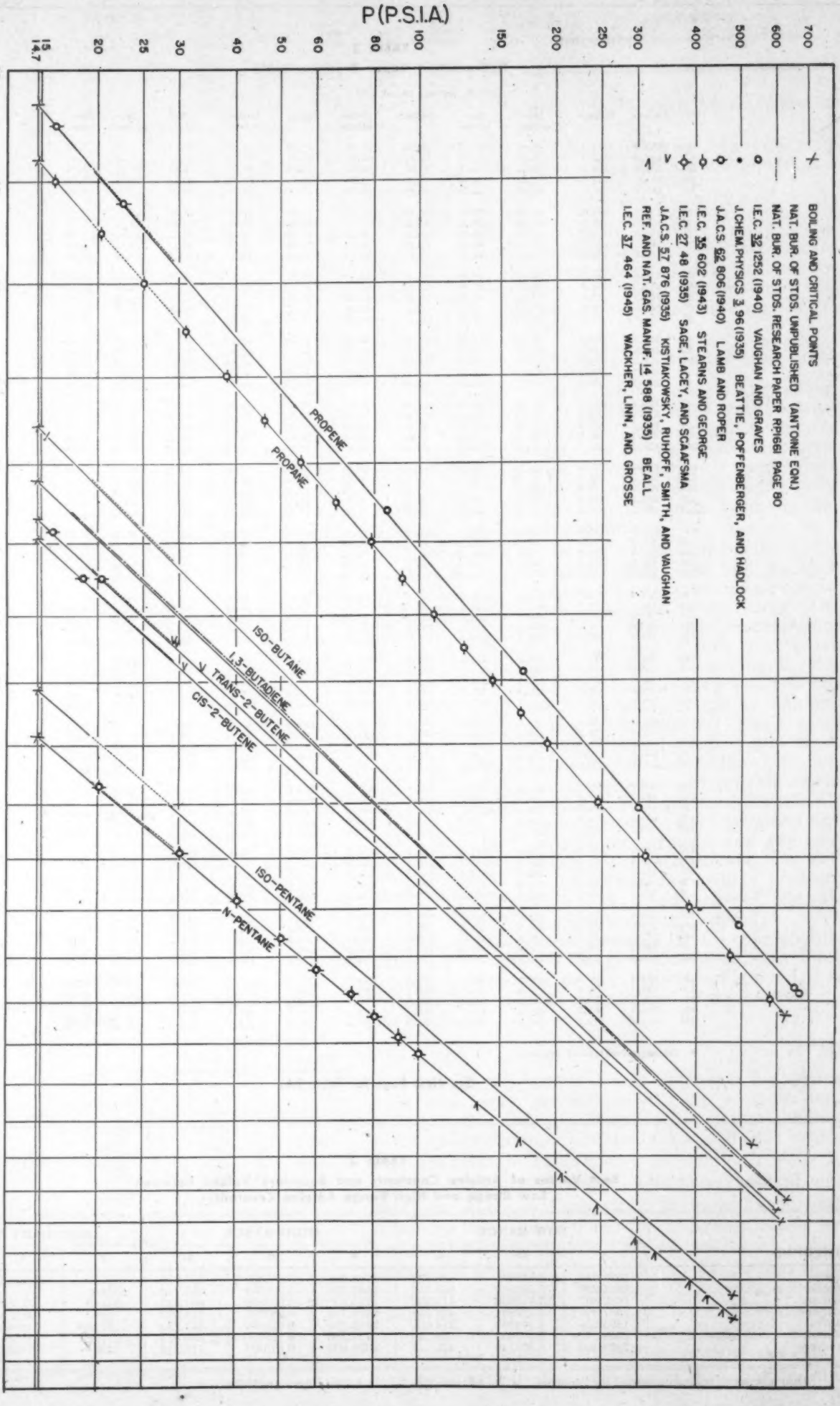
P(P.S.I.A)
T (°F)
PROPENE
PROPANE
ISO-BUTANE
1,3-BUTADIENE
TRANS-2-BUTENE
CIS-2-BUTENE
ISO-PENTANE
N-PENTANE
BOILING AND CRITICAL POINTS
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FIGURE 1

See article "High Range Vapor Pressure and Antoine Equation Constants of some C_2 to C_6 Hydrocarbons" which starts on the opposite page.

TABLE 2

Hydrocarbon Vapor Pressure Table

(Lb. per sq. in. at t°F)

t°F	Propene	Propane	Isobutane	Isobutene	1-Butene	1,3-Butadiene	Butane	trans-2 Butene	cis-2 Butene	Isopentane	Pentane
60	130.8	107.3	37.71	32.32	31.54	29.77	25.78	24.79	22.45	9.58	7.12
62	134.8	110.7	39.04	33.50	32.69	30.87	26.74	25.73	23.31	9.97	7.47
64	138.8	114.1	40.41	34.71	33.87	32.01	27.73	26.71	24.21	10.38	7.74
66	143.0	117.6	41.81	35.96	35.08	33.18	28.74	27.71	25.13	10.80	8.06
68	147.2	121.1	43.25	37.24	36.33	34.38	29.80	28.74	26.08	11.23	8.40
70	151.5	124.8	44.72	38.56	37.62	35.61	30.87	29.80	27.06	11.67	8.75
72	156.0	128.5	46.24	39.91	38.94	36.89	31.98	30.90	28.07	12.14	9.11
74	160.5	132.3	47.79	41.29	40.29	38.19	33.12	32.02	29.11	12.61	9.48
76	165.1	136.2	49.39	42.72	41.68	39.53	34.28	33.17	30.17	13.10	9.86
78	169.8	140.2	51.02	44.19	43.11	40.92	35.48	34.37	31.27	13.60	10.26
80	174.6	144.2	52.70	45.69	44.58	42.33	36.71	35.59	32.40	14.12	10.67
82	179.5	148.4	54.42	47.24	46.08	43.79	37.99	36.84	33.55	14.66	11.10
84	184.5	152.6	56.18	48.82	47.63	45.28	39.29	38.13	34.75	15.21	11.53
86	189.6	156.9	57.98	50.45	49.21	46.81	40.63	39.47	35.98	15.78	11.98
88	194.8	161.4	59.83	52.11	50.83	48.38	41.99	40.83	37.24	16.37	12.44
90	200.2	165.9	61.73	53.83	52.50	49.99	43.40	42.23	38.53	16.97	12.92
92	205.6	170.5	63.67	55.58	54.21	51.65	44.84	43.66	39.86	17.60	13.42
94	211.1	175.2	65.66	57.37	55.96	53.34	46.32	45.14	41.23	18.24	13.93
96	216.7	179.9	67.69	59.21	57.75	55.08	47.84	46.65	42.63	18.89	14.45
98	222.5	184.8	69.77	61.09	59.59	56.86	49.39	48.20	44.07	19.57	14.99
100	228.3	189.8	71.90	63.02	61.48	58.70	50.99	49.80	45.54	20.26	15.55
102	234.3	194.9	74.07	64.99	63.40	60.57	52.61	51.42	47.06	20.98	16.12
104	240.3	200.0	76.30	67.02	65.37	62.48	54.29	53.10	48.61	21.71	16.71
106	246.5	205.3	78.58	69.09	67.40	64.46	56.01	54.82	50.21	22.47	17.31
108	252.8	210.7	80.91	71.21	69.47	66.47	57.77	56.58	51.85	23.24	17.94
110	259.2	216.1	83.29	73.39	71.58	68.53	59.56	58.38	53.53	24.04	18.58
112	265.8	221.7	85.73	75.61	73.75	70.62	61.42	60.23	55.24	24.86	19.24
114	272.4	227.4	88.22	77.88	75.96	72.78	63.30	62.12	57.01	25.70	19.92
116	279.2	233.2	90.76	80.21	78.23	74.99	65.23	64.05	28.81	26.56	20.61
118	286.1	239.0	93.36	82.60	80.55	77.25	67.20	66.05	60.67	27.45	21.33
120	293.0	245.0	96.01	85.01	82.91	79.56	69.22	68.07	62.55	28.36	22.07
122	300.2	251.1	98.72	87.51	85.34	81.93	71.30	70.15	64.49	29.29	22.82
124	307.4	257.3	101.5	90.04	87.81	84.34	73.41	72.28	66.48	30.25	23.60
126	314.8	263.6	104.3	92.64	90.34	86.81	75.56	74.45	68.51	31.22	24.40
128	322.3	270.1	107.2	95.28	92.93	89.34	77.77	76.68	70.58	32.23	25.22
130	329.9	276.6	110.1	98.00	95.56	91.91	80.03	78.96	72.71	33.26	26.06
132	337.7	283.2	113.1	100.8	98.26	94.56	82.32	81.28	74.89	34.31	26.92
134	345.6	290.0	116.2	103.6	101.0	97.25	84.68	83.67	77.11	35.39	27.80
136	353.6	296.9	119.3	106.5	103.8	100.0	87.09	86.10	79.40	36.51	28.72
138	361.7	303.9	122.5	109.4	106.7	102.8	89.55	88.59	81.72	37.64	29.65
140	369.9	311.0	125.8	112.4	109.6	105.7	92.05	91.13	84.10	38.80	30.61
142	378.3	318.3	129.1	115.5	112.6	108.6	94.63	93.73	86.55	39.99	31.58
144	386.8	325.6	132.5	118.6	115.7	111.6	97.26	96.38	89.03	41.21	32.58
146	395.5	333.0	135.9	121.8	118.8	114.7	99.92	99.09	91.57	42.46	33.61
148	404.4	340.6	139.4	125.1	122.0	117.8	102.7	101.9	94.16	43.73	34.66
150	413.3	348.4	143.0	128.4	125.2	121.0	105.5	104.7	96.81	45.04	35.75
152	422.4	356.2	146.6	131.8	128.5	124.2	108.3	107.6	99.52	46.38	36.85
154	431.6	364.2	150.3	135.2	131.9	127.5	111.2	110.5	102.3	47.74	37.99
156	440.9	372.2	154.1	138.7	135.3	130.9	114.1	113.5	105.1	49.14	39.15
158	450.4	380.4	158.0	142.3	138.8	134.3	117.2	116.6	108.0	50.56	40.34
160	460.1	388.7	161.9	146.0	142.4	137.9	120.2	119.7	110.9	52.03	41.55
162	469.8	397.2	165.9	149.7	146.0	141.4	123.4	122.9	113.9	53.52	42.80
164	479.8	405.8	170.0	153.5	149.7	145.1	126.6	126.2	117.0	55.05	44.07
166	489.8	414.5	174.1	157.4	153.5	148.8	129.8	129.5	120.1	56.61	45.37
168	500.0	423.3	178.3	161.3	157.3	152.6	133.1	132.9	123.3	58.19	46.71
170	510.4	432.2	182.6	165.3	161.2	156.5	136.5	136.3	126.6	59.83	48.06
172	520.9	441.3	186.9	169.4	165.2	160.4	140.0	139.9	129.9	61.48	49.45
174	531.5	450.6	191.4	173.6	169.3	164.4	143.5	143.5	133.3	63.19	50.88
176	542.3	460.0	195.9	177.8	173.4	168.5	147.1	147.1	136.7	64.91	52.34
178	553.3	469.5	200.5	182.1	177.6	172.6	150.7	150.8	140.2	66.69	53.83
180	564.4	479.1	205.2	186.5	181.9	176.8	154.4	154.6	143.8	68.49	55.35
182	575.6	488.9	209.9	191.0	186.2	181.2	158.2	158.5	147.5	70.34	56.90
184	587.0	498.9	214.7	195.5	190.6	185.5	162.0	162.5	151.2	72.22	58.49
186	598.6	508.9	219.6	200.2	195.1	190.0	166.0	166.5	155.0	74.14	60.11
188	610.3	519.1	224.6	204.9	199.7	194.5	170.0	170.6	158.9	76.09	61.77
190	622.1	529.4	229.7	209.7	204.4	199.2	174.0	174.7	162.8	78.09	63.46
192	634.1	539.9	234.8	214.5	209.1	203.9	178.1	179.0	166.8	80.13	65.19
194	646.3	550.5	240.0	219.5	214.0	208.7	182.3	183.3	170.9	82.21	66.95
196	658.7	561.2	245.4	224.5	218.9	213.5	186.6	187.7	175.1	84.32	68.75
198	671.2[a]*	572.1	250.8	229.6	223.8	218.4	190.9	192.1	179.3	86.48	70.58

* Indicates Position of Critical

(See Next Page for Table 2A)

TABLE 3

Best Values of Antoine Constants and Boundary Values Between Low Range and High Range Antoine Constants

COMPOUND	LOW RANGE			HIGH RANGE			BOUNDARY VALUES		
	A	B	C	A	B	C	°F	T_r	P_r
Propene	1528.314	5.243591	428.790	2194.298	5.850499	527.719	61.3[b]	0.793	0.201
Propane[a]	1842.226	5.532988	465.710	1842.226	5.532988	465.710	None	.....	.....
Isobutane	1689.897	5.145680	413.882	2245.800	5.653856	491.523	99.99[b]	0.762	0.136
Isobutene	1957.028	5.427894	439.754	2878.318	6.082497	574.833	167[c]	0.835	0.279
1-Butene	1493.254	4.944063	374.651	2155.070	5.611607	461.232	182.69[b]	0.849	0.313
n-Butane	2208.373	5.607685	466.240	3727.356	6.520950	680.577	177.3[c]	0.834	0.275
n-Pentene	2168.407	5.361146	420.117	4323.000	6.615431	713.312	270[c]	0.863	0.344

[a] Only one set of constants needed for Propane. [b] From plot. [c] Actual Literature Value.

TABLE 2A

Hydrocarbon Vapor Pressure Table

(Lb. per sq. in. at t°F)

t°F	Propene	Propane	Isobutane	Isobutene	1-Butene	1,3-Butadiene	Butane	trans-2 Butene	cis-2 Butene	Isopentane	Pentane
200	683.8	583.1	256.2	234.8	228.9	223.5	195.3	196.7	183.6	88.68	72.46
202	696.6	594.3	261.8	240.1	234.1	228.6	199.8	201.3	187.9	90.93	74.37
204	709.5	605.7	267.5	245.4	239.3	233.8	204.4	206.0	192.4	93.21	76.32
206	722.6	617.1[c]	273.2	250.9	244.6	239.1	209.0	210.8	197.0	95.54	78.31
208	736.0	628.7	279.0	256.5	250.0	244.4	213.7	215.7	201.6	97.91	80.33
210	749.5	640.5	285.0	262.1	255.5	249.9	218.5	220.6	206.3	100.3	82.40
212	763.1	652.4	291.0	267.8	261.1	255.4	223.4	225.7	211.0	102.8	84.51
214	776.9	664.5	297.1	273.6	266.7	261.0	228.3	230.8	215.9	105.3	86.66
216	790.8	676.7	303.2	279.5	272.5	266.8	233.4	236.0	220.8	107.8	88.85
218	804.9	689.1	309.5	285.5	278.3	272.6	238.5	241.3	225.8	110.4	91.09
220	819.2	701.6	315.9	291.6	284.2	278.5	243.7	246.7	231.0	113.1	93.36
222	833.6	714.2	322.4	297.8	290.3	284.5	248.9	252.1	236.1	115.8	95.68
224	848.3	727.1	328.9	304.0	296.4	290.6	254.3	257.6	241.4	118.5	98.04
226	863.1	740.0	335.6	310.4	302.6	296.8	259.8	263.3	246.8	121.3	100.4
228	878.1	753.1	342.3	316.9	308.9	303.1	265.3	269.0	252.2	124.1	102.9
230	893.2	766.4	349.2	323.4	315.3	309.5	270.9	274.8	257.7	127.0	105.4
232	908.5	779.9	356.1	330.1	321.7	315.9	276.6	280.7	263.4	130.0	107.9
234	924.0	793.5	363.1	336.9	328.3	322.5	282.4	286.7	269.1	133.0	110.5
236	939.6	807.2	370.3	343.7	335.0	329.1	288.2	292.8	274.9	136.0	113.2
238	955.4	821.1	377.5	350.7	341.8	335.9	294.2	299.0	280.8	139.1	115.8
240	971.4	835.2	384.8	357.7	348.6	342.8	300.2	305.3	286.8	142.2	118.6
242	987.7	849.4	392.3	364.8	355.6	349.7	306.3	311.7	292.9	145.4	121.4
244	1004	863.8	399.8	372.1	362.7	356.8	312.6	318.1	299.0	148.7	124.2
246	1020	878.3	407.4	379.5	369.8	364.0	318.9	324.7	305.3	152.0	127.1
248	1037	893.0	415.1	386.9	377.1	371.3	325.3	331.4	311.7	155.3	130.0
250	1054	907.9	423.0	394.4	384.4	378.7	331.8	338.1	318.1	158.7	133.0
252			430.9	402.1	391.9	386.2	338.3	345.0	324.7	162.2	136.0
254			438.9	409.8	399.5	393.8	345.0	351.9	331.3	165.7	139.1
256			447.1	417.7	407.1	401.4	351.7	359.0	338.0	169.3	142.2
258			455.3	425.7	414.9	409.2	358.6	366.1	344.9	172.9	145.4
260			463.7	433.8	422.8	417.1	365.5	373.4	351.8	176.6	148.6
262			472.1	442.0	430.8	425.1	372.6	380.8	358.9	180.4	151.9
264			480.7	450.3	438.8	433.2	379.7	388.3	366.0	184.2	155.2
266			489.4	458.7	447.0	441.4	387.0	395.8	373.3	188.0	158.6
268			498.1	467.2	455.3	449.7	394.3	403.5	380.6	191.9	162.1
270			507.0	475.8	463.7	458.2	401.8	411.3	388.1	195.9	165.6
272			516.0	484.5	472.2	466.7	409.3	419.2	395.6	200.0	169.1
274			525.1	493.3	480.9	475.4	417.0	427.2	403.3	204.1	172.8
276			534.3[c]	502.3	489.6	484.2	424.7	435.3	411.0	208.2	176.5
278			543.7	511.4	498.4	493.1	432.5	443.5	418.9	212.5	180.2
280			553.1	520.6	507.4	502.1	440.5	451.8	426.9	216.8	184.0
282			562.6	529.9	516.4	511.2	448.5	460.3	435.0	221.1	187.8
284			572.3	539.3	525.6	520.5	456.7	468.8	443.2	225.5	191.7
286			582.1	548.9	534.9	529.8	464.9	477.5	451.5	230.0	195.7
288			591.9	558.5	544.3	539.3	473.2	486.2	459.9	234.5	199.7
290			601.9	568.3	553.8	548.8	481.7	495.1	468.4	239.1	203.8
292			612.1	578.1	563.4	558.6	490.2	504.1	477.0	243.8	208.0
294			622.3	588.2[c]	573.2	568.4	498.9	513.2	485.8	248.6	212.2
296			632.6	598.3	583.0	578.4	507.7	522.4	494.7	253.4	216.5
298			643.1	608.6	593.0[c]	588.4	516.5	531.8	503.6	258.2	220.8
300			653.6	618.9	603.1	598.6	525.5	541.2	512.7	263.2	225.2
302			664.3	629.4	613.3	608.8	534.6	550.7	521.9	268.2	229.7
304			675.1	639.9	623.6	619.2	543.8	560.4	531.3	273.2	234.2
306			686.1	650.7	634.1	629.8[c]	553.1[c]	570.3	540.7	278.4	238.8
308			697.1	661.5	644.6	640.5	562.4	580.2	550.2	283.6	243.5
310			708.3	672.5	655.3	651.3	572.0	590.2	559.9	288.9	248.2
312			719.5	683.5	666.1	662.2	581.7	600.4[c]	569.7	294.2	253.0
314			731.0	694.8	677.0	673.3	591.4	610.7	579.6	299.7	257.9
316			742.5	706.1	688.0	684.5	601.3	621.2	589.7	305.2	262.8
318			754.1	717.6	699.2	695.7	611.2	631.7	599.8	310.7	267.8
320			765.9	729.2	710.5	707.1	621.3	642.3	610.0[c]	316.4	272.9
322			777.8	740.9	721.9	718.7	631.5	653.1	620.4	322.1	278.0
324			789.8	752.7	733.5	730.3	641.7	664.0	630.9	327.9	283.2
326			801.9	764.7	745.1	742.1	652.1	675.1	641.6	333.7	288.5
328			814.1	776.8	756.9	754.1	662.7	686.2	652.4	339.6	293.8
330			826.6	789.0	768.8	766.2	673.5	697.5	663.2	345.7	299.3

desirable because of economy of time and effort.

Thomson[8] and others[11] have shown that the Antoine equation,

$$\log P = B - A/(T + C) \qquad (1)$$

is capable of representing vapor pressure data very accurately over limited ranges. The recent correlation of C_4 hydrocarbon vapor pressures noted earlier[5] utilizes the relation,

$$\log P = B - A/T + \Delta \qquad (1A)$$

over the entire range. In the above, A, B, and C are empirical constants evaluated from vapor pressure data, and Δ is a logarithmic pressure or reciprocal temperature *residual,* which depends upon a number of factors in a generally unknown manner.

In the present correlations, the normal boiling point and the critical are used as reference points for high range vapor pressures. The Bureau of Standards' recent values of the normal boiling points are employed for all compounds. For the criticals, Cragoe's selected values[3] are used, except as follows: Later values were available for isobutane,[2] for 1-butene,[2, 6] and for 1, 3-butadiene.[7] Values that give better agreement with reliable vapor pressure data are used for the C_3 hydrocarbons: For propane, the critical temperature and pressure given by Deschner and Brown.[4] For propene, the critical pressure by Vaughan and Graves.[9] The values are given in Table 1.

Plots of recent vapor pressure data on the coordinates log P and 1/T°R. give almost straight lines from 1 atmosphere to the critical. (Figure 1.) A straight line on these coordinates is represented by the relation,

$$\log P = B - A/T \qquad (2)$$

or Equation 1A with $\Delta = O$. Equation 2 requires only two vapor pressure points for a compound, which in this case are taken as the normal boiling point and the critical. Constants A and B thus evaluated are given in Table 1.

Vapor pressures calculated with these constants are given in Table 2. Comparing these with the best literature values, it is found that the table values for pressures average about 0.9 percent low in the region around 50 psia, and average about 1.2 percent high in the region around 300 psia. The corresponding temperature deviations are about —0.6° F. in the low range and about +1.0° F. in the high range. Comparison with the values of Hachmuth, et al[5] shows excellent agreement in the lower part of the range and the

TABLE 4

Average Temperature Deviations from Best Literature Values, in °F.

	Table 1	Table 3 (Antoine Eqs.)	Thomson (Antoine Eqs.)
LOW RANGE			
Propene	0.251	0.046	.....
Propane[a]	0.461	0.320	0.762
Isobutane	0.168	0.134	.....
Isobutene	0.338	0.207	0.137[b]
1-Butene	0.827	0.499	.....
n-Butane	0.232	0.107	0.722
n-Pentane	0.635	0.116	0.273
HIGH RANGE			
Propene	0.597	0.271	.....
Propane[a]	0.461	0.320	0.762
Isobutane	0.569	0.228	.....
Isobutene	0.590	0.114	0.137
1-Butene	0.545	0.450	.....
n-Butane	1.017	0.188	.....
n-Pentane	0.869	0.430	.....

[a] Only one set of constants needed for Propane.
[b] Equation covers part of High Range as well as the Low Range.

present values to be about 1 psi higher in the higher part of the range. The vapor pressures of n-butane and of trans-2-butene cross at about 175° F. according to both correlations. The values of Table 2 are evidently interconsistent and may be used with confidence for column plate calculations.

Temperature deviations between Table 2 and the best literature values are plotted against reduced temperature on Figures 2, 3, 4, 5, 6, 7, and 8, which include only those hydrocarbons for which there are adequate experimental data in the high range. With the exception of propane, the residuals are negative in the lower part of the range, they cross zero and have positive values at the higher temperatures. Thus when A and B are evaluated from the normal boiling point and critical of a compound, the equation is exact at the two reference points and also at some intermediate point. This suggests that when a more accurate fit is desired, Equation 1 may be applied to a compound using one set of constants evaluated for the low pressure range, and another set of constants evaluated for the high pressure range. The junction of the equations may then be taken at the point of zero deviation of Equation 2, which for six of the seven hydrocarbons of the Figures lies at some value of reduced temperature between 0.75 and 0.9. After extensive study of methods of evaluating Antoine equation constants, Thomson recommended a procedure of using two equations between the normal boiling point and the critical, but with the junction at a constant reduced temperature of 0.8.

Evaluation of constants A, B, and C in Equation 1 requires three vapor pressure points: For the low range, these are the normal boiling point, the junction or point of zero deviation for Equation 2, and an intermediate point, preferably where the deviation for Equation 2 is near its maximum. The same considerations apply for evaluation of the high range constants, using the critical, the junction, and an intermediate point. This procedure requires determination of the point of zero deviation, which may be done by

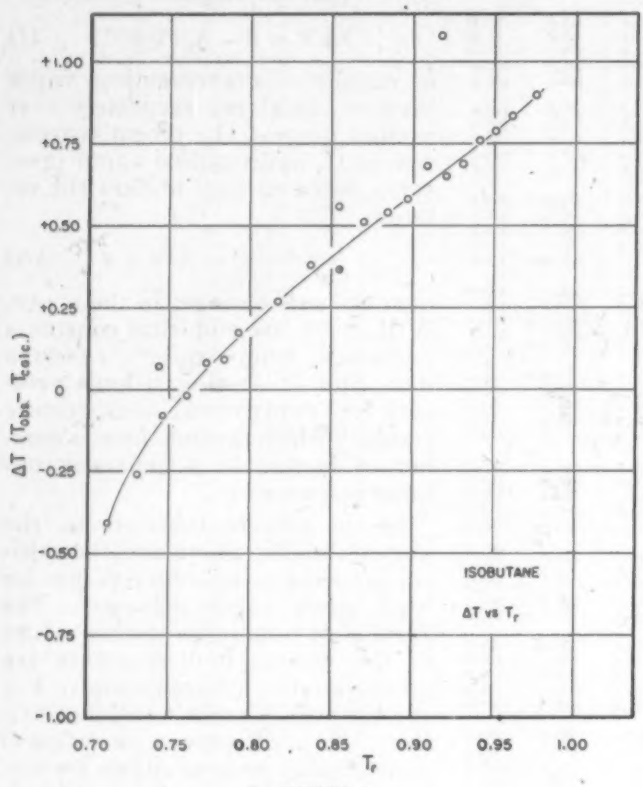

FIGURE 4

Temperature residuals for isobutane, from Equation 2.

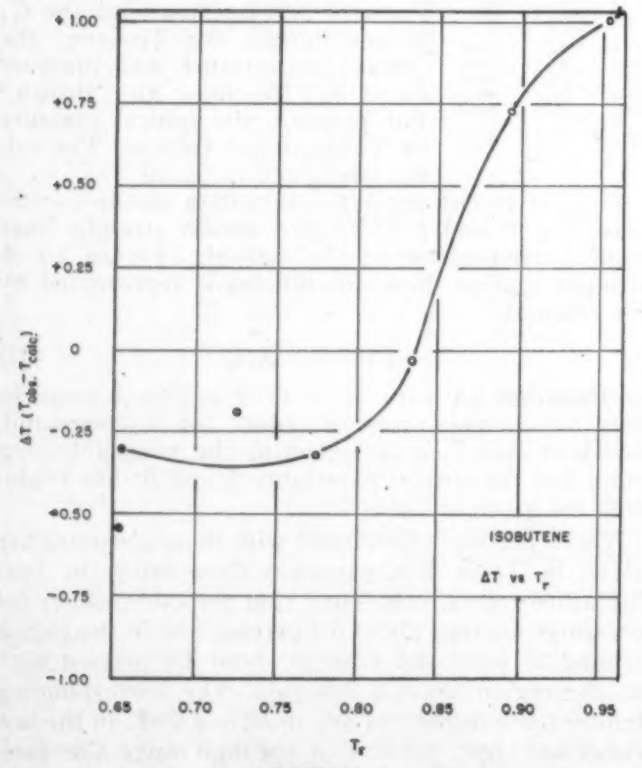

FIGURE 5

Temperature residuals for isobutene, from Equation 2.

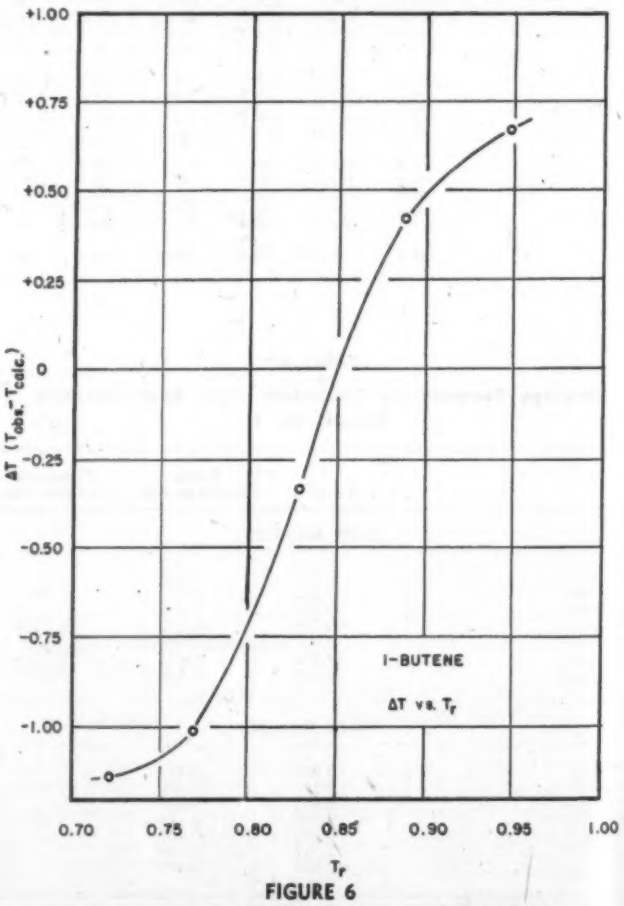

FIGURE 6

Temperature residuals for i-butene, from Equation 2.

plotting experimental data as on Figure 1. Where a dependable experimental value is close to the point of zero deviation, it may be used as the junction instead of the actual zero point, with negligible difference in the accuracy of fit.

Low- and high range Antoine equation constants along with the junction values are given in Table 3 for the seven hydrocarbons of the Figures. These constants were obtained by simultaneous solution of Equation 1 at three points as noted. The average of temperature deviations between Equation 1 using the constants of Table 3 and the literature values are given in Table 4. Temperature deviations using Thomson's Antoine constants are included for comparison.

It is seen that the present procedure gives a better fit of the equation to the data than was obtained by taking the junction at a constant reduced temperature of 0.8 for all compounds. Figure 9 is a plot of reduced pressure against reduced temperature, where P_r and T_r are taken at the intermediate point of zero residual. For six of the seven hydrocarbons, the points form a common straight line. This relation is probably sufficiently general to be helpful in extrapolating and interpolating hydrocarbon vapor pressures in the superatmospheric range.

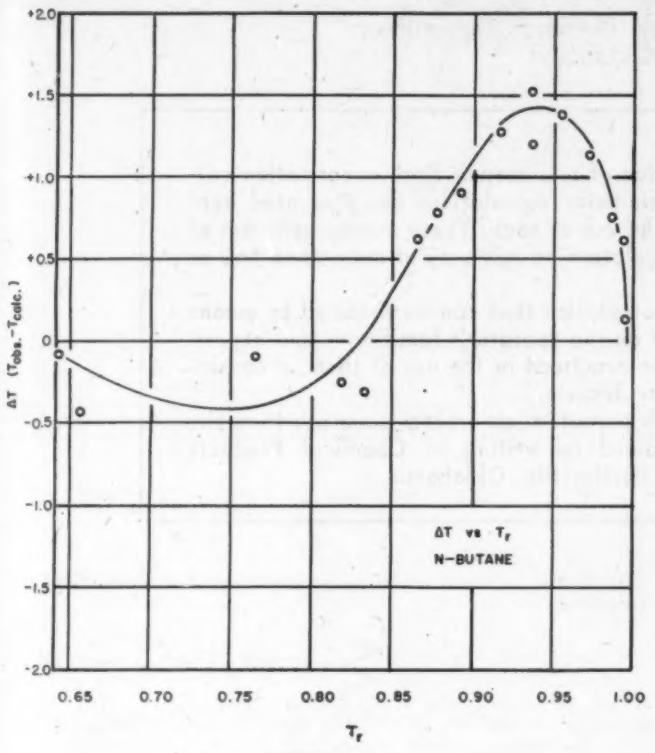

FIGURE 7
Temperature residuals for n-butane, from Equation 2.

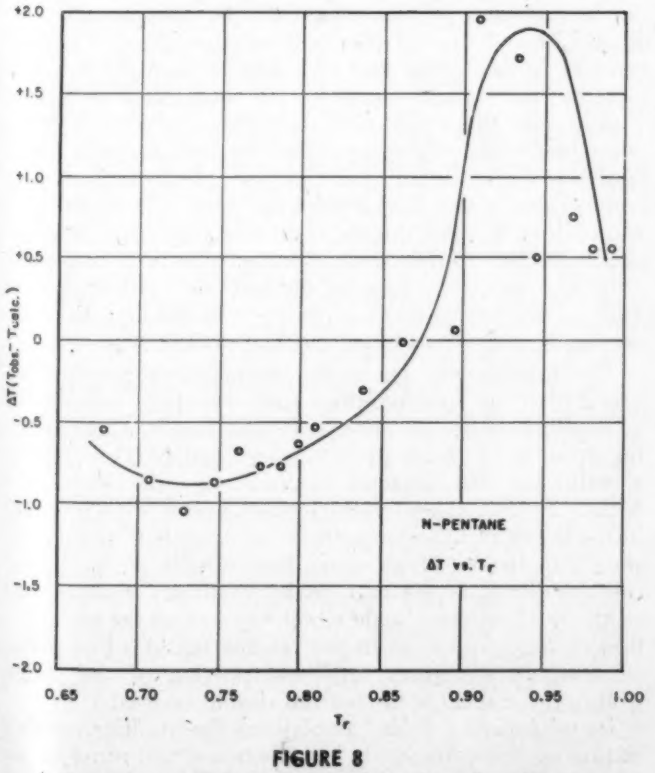

FIGURE 8
Temperature residuals for n-pentane, from Equation 2.

ACKNOWLEDGMENT

The preparation of Table 2 and the work leading to it were done under sponsorship of RFC Office of Rubber Reserve in connection with the Government synthetic rubber program.

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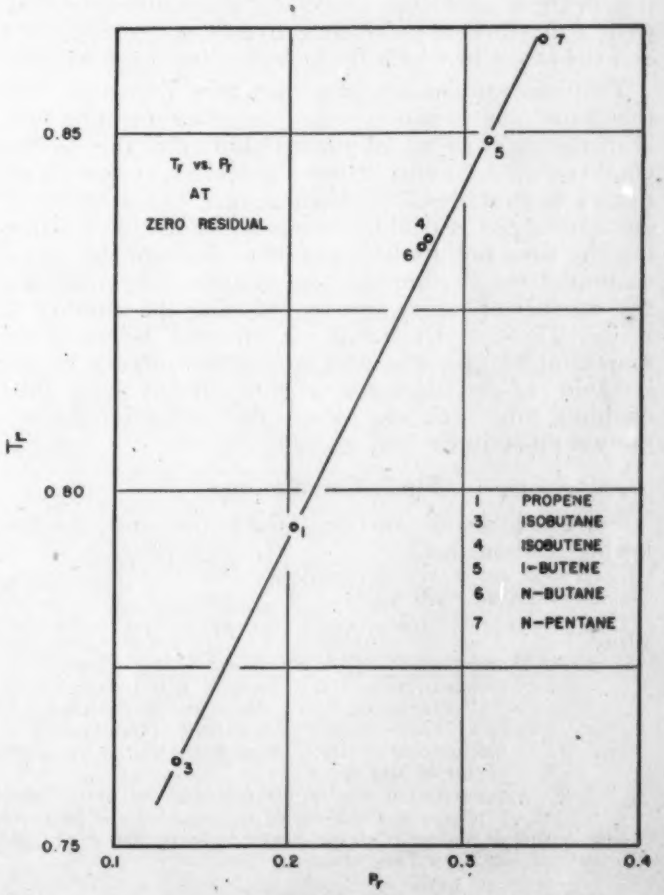


FIGURE 9

Nomographs for Vapor-Liquid Equilibrium Calculations

FRED H. POETTMANN and R. B. RICE
Phillips Petroleum Company, Research Department,
Bartlesville, Oklahoma

TWO nomographs for the graphical solution of equilibrium flash vaporization calculations, dew-point calculations and bubble-point calculations are presented here and numerical examples cited to describe the use of each. These nomographs are of particular help in reducing the time of calculation on mixtures of more than five or six components.

The authors point out that the speed of calculation that can be obtained by means of these nomographs is naturally dependent on the operator's familiarity and experience with them. "Once a person has become practiced in the use of them, a considerable saving in time can be effected," they declare.

Full-scale copies of the nomographs which permit much greater accuracy than the accompanying reproductions may be obtained by writing to Chemical Products Department, Phillips Petroleum Company, Bartlesville, Oklahoma.

PETROLEUM engineers spend a great deal of time performing tedious trial-and-error calculations, a large part of which are flash calculations, dew-point calculations and bubble-point calculations. Hence it is desirable to employ graphical methods in the solution of these problems, especially when one is dealing with a mixture of more than five or six components and the range in which the solution lies is not known.

Two nomographs are presented here for facilitating the above mentioned types of calculations. The first is in the form of an alignment chart and the second employs labeled contour lines. Each has its own particular advantages, but both reduce the amount of labor involved in trial-and-error solutions by shortening the time required for one trial through the elimination of the arithmetic computations, by reducing the number of errors and by reducing the number of trials. The latter savings is effected because the nomographs give the user a graphic picture of the behavior of the fundamental function involved, thus enabling him to choose an assumed value for the unknown more judiciously.

Flash Calculations

For 100 mols of a mixture of n components, the following relations hold:

$$V + L = 100 \quad (1)$$

$$Vy_i + Lx_i = 100\, x_{ei}, \; i = 1, 2, \ldots, n \quad (2)$$

$$y_i = K_i x_i, \; i = 1, 2, \ldots, n \quad (3)$$

where

V = mol percent of the mixture in vapor phase,
L = mol percent of the mixture in liquid phase,
y_i = mol percent of the i^{th} component in the vapor,
x_i = mol percent of the i^{th} component in the liquid,
x_{ei} = mol percent of the i^{th} component in the composite (liquid and vapor)
K_i = vaporization equilibrium constant for the i^{th} component at the assumed temperature and pressure.

Eliminating V and y_i from equation (2) by means of (1) and (3) and solving for x_i, one obtains

$$x_i = \frac{100\, x_{ei}}{L + (100 - L)\, K_i}, \; i = 1, 2, \ldots, n \quad (4)$$

This is the relation solved by each nomograph. Once the value for which

$$\sum_{i=1}^{n} x_i = 100 \quad (5)$$

is found, V and y_i are readily computed from relations (1) and (2), respectively.

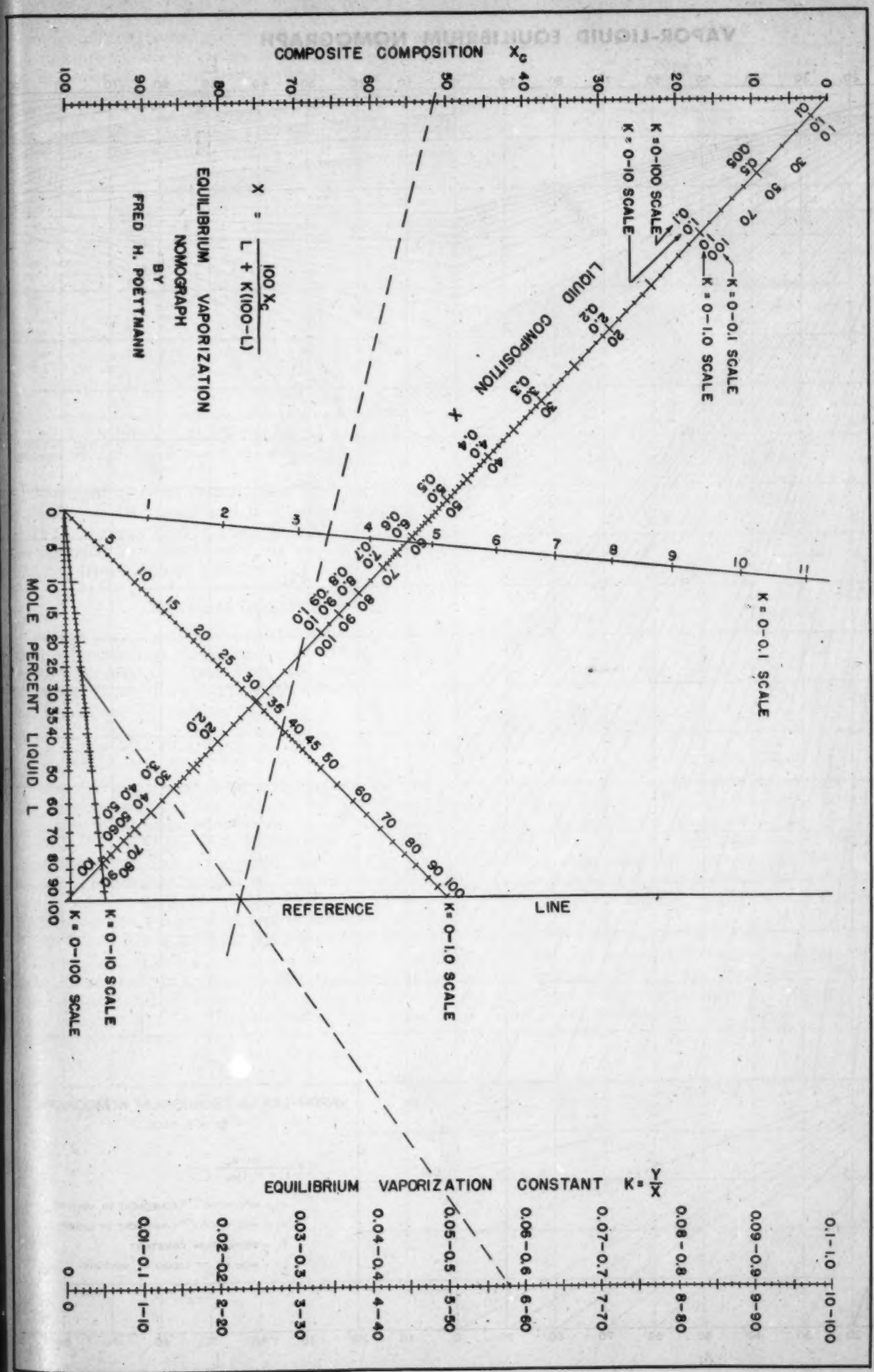
To illustrate the use of the first nomograph in performing flash calculations, let us assume that a mixture containing 51.5 mol percent propane is flashed at one atmosphere pressure and 45° F. also assume that K = 5.79 and that of the total mixture 26 mol percent remains in the liquid state. To find the mol percent propane in the residual liquid by the nomograph of Figure 1 note that there are four L scales radiating from a point just to the left of center at the bottom, each corresponding to the range of K indicated. Following the key connect 26 on the K = 0-10 scale with 5.79 on the corresponding K scale on the right and mark the intersection with the reference line. Connect this reference point with 51.5 on the x_e scale on the left and read a value of 11.2 for the mol percent of propane in the liquid residue on the K = 0-10 scale of the diagonal x scale.

To illustrate the use of the second nomograph (Figure 2) let us assume that for a certain component, $x_e = 40$, K = 1.5 and that L is taken as 50. Then starting from the L scale in the upper right-hand section at a value of 50, descend vertically to the curve for K = 1.5. (For cases in which the L and K lines intersect in the lower right-hand portion, the L scale at the bottom should be used.) Next, move horizontally to the left to the line for $x_e = 40$, and finally vertically upward to a value of 32 on the x scale which represents the mol fraction of the component in the residue liquid. (For points occuring in the lower left-hand portion of the nomograph, the x scale at the bottom should be used.)

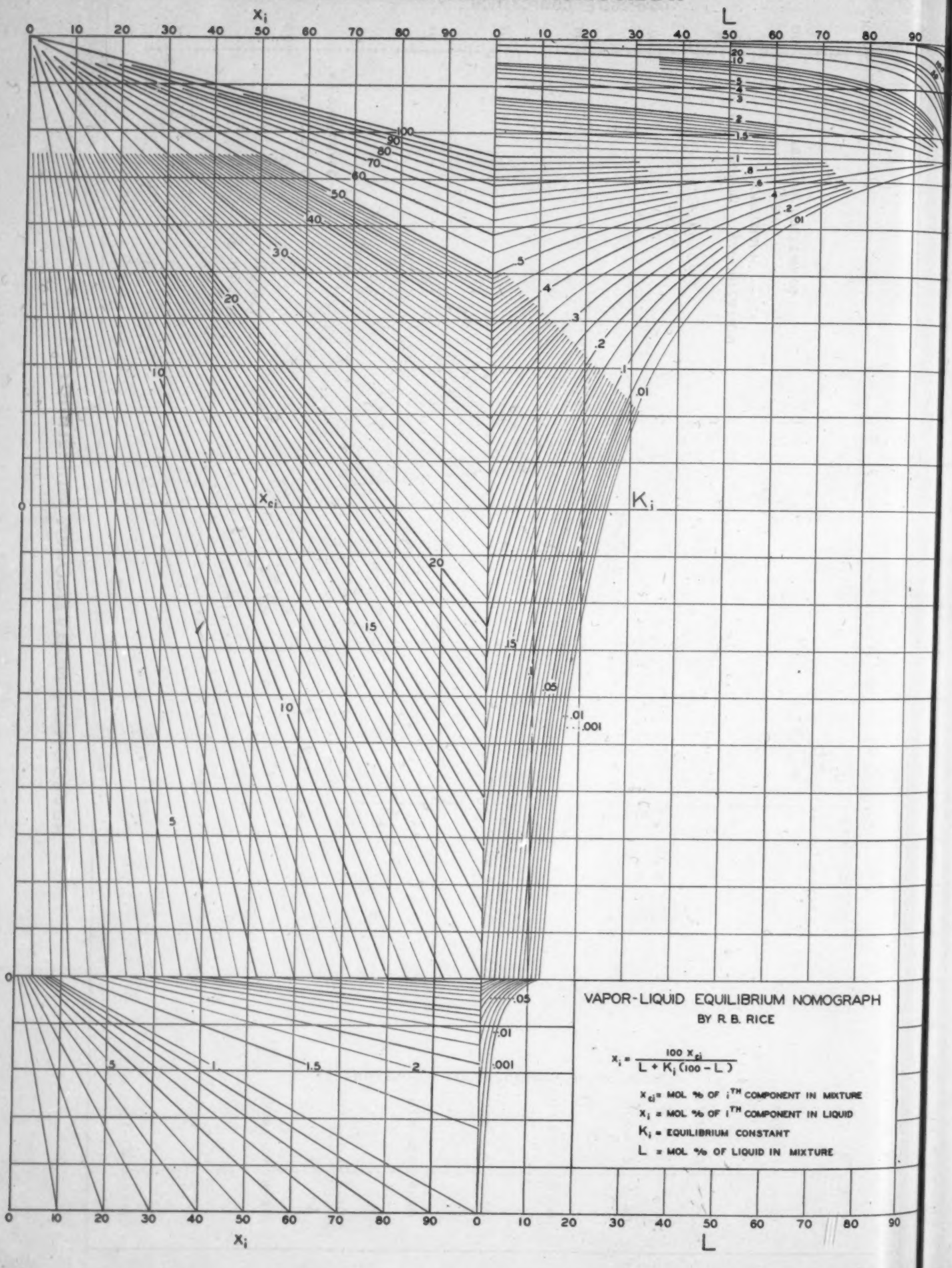
In performing flash calculations for multicomponent mixtures, the value of L is not known and must be assumed. Then the operation described above for each

MOLE PERCENT LIQUID L

Dec


EQUILIBRIUM VAPORIZATION NOMOGRAPH
BY
FRED H. POETTMANN
EQUILIBRIUM VAPORIZATION NOMOGRAPH
$X = \frac{100\,X_c}{L + K(100-L)}$
COMPOSITE COMPOSITION X_c
LIQUID COMPOSITION X
MOLE PERCENT LIQUID L
EQUILIBRIUM VAPORIZATION CONSTANT $K = \frac{Y}{X}$
K = 0-100 SCALE
K = 0-10 SCALE
K = 0-1.0 SCALE
K = 0-0.1 SCALE
REFERENCE LINE

VAPOR-LIQUID EQUILIBRIUM NOMOGRAPH


X_i
L
X_{ci}
K_i
0 10 20 30 40 50 60 70 80 90
100 90 80 70 60 50 40 30 20 10
20 10 5 4 3 2 1.5 1 .8 .6 .4 .2 .01
5 4 3 .2 .1 .01
20 15 10 5
15 10 5 1 .5 .1 .05 .01 .001
.05 .01 .001
.5 1 1.5 2
VAPOR-LIQUID EQUILIBRIUM NOMOGRAPH
BY R. B. RICE
$x_i = \frac{100\, x_{ci}}{L + K_i(100 - L)}$
x_{ci} = MOL % OF i^{TH} COMPONENT IN MIXTURE
x_i = MOL % OF i^{TH} COMPONENT IN LIQUID
K_i = EQUILIBRIUM CONSTANT
L = MOL % OF LIQUID IN MIXTURE

nomograph is repeated for each of the n components in the mixture and the x_i values added to determine whether or not relation (5) has been satisfied. If not, then another value of L is assumed and the process repeated.

Example 1. In order that the reader may check his use of the nomographs, the Table 1 includes a summary of a calculation for a mixture flashed at 20° F. and 450 psia.

TABLE 1

Component	x_{ci}	K_i*	Assume L = 20 x_i	Assume L = 30 x_i	Assume L = 25 x_i
N_2	3.20	28.0	0.1	0.2	0.1
CH_4	57.75	4.9	14.0	15.5	14.5
C_2H_6	16.47	.76	20.4	19.8	20.0
C_3H_8	11.49	.24	29.3	24.5	26.6
C_4H_{10}	6.55	.072	25.4	18.7	21.4
C_5H_{12}	2.73	.021	12.6	8.7	10.3
C_6H_{14}	1.26	.006	6.1	4.1	4.9
C_7H_{16}+	0.55	.002	2.7	1.8	2.2
Total	100.00	....	110.6 L too small	93.3 L too large	100.0 L correct

* All K constants used in Examples 1 and 2 were calculated from fugacity coefficients published by Gamson and Watson, National Petroleum News, Technical Section, September 6, 1944.

It should be noted that in cases in which the value of x_c is below 10 percent and K is quite large (as in the case of nitrogen above), more accurate values of x can be obtained by multiplying x_c by ten and then dividing the resultant value of x by ten.

Dew-Point Calculations

In the case of dew-point calculations it is desired to find a temperature and pressure at which the first infinitesimal drop of liquid condenses out of a given gas mixture. Hence, L = O and, upon substitution of y_i for x_{ci}, equation (4) reduces to

$$x_i = \frac{y_i}{K_i}, \quad i = 1, 2, \ldots, n, \tag{3'}$$

In order to perform these computations by means of the nomographs, one finds the set of K_i values corresponding to the assumed temperature and pressure. Then one takes L = O and goes through the same steps outlined above for each nomograph, remembering the x_c is now y. The process is repeated for each component and the x_i values added. If relation (5) has not been satisfied, then a new temperature and/or pressure (and hence a new set of K's is assumed and the process repeated.

Example 2. Table 2 contains a summary of a dew-point calculation on a gas with the same composition as the mixture used in Example 1. In this case the pressure has been fixed at 450 psia and the dew-point temperature calculated.

TABLE 2

Component	x_{ci} ($=y_i$)	Assume t = 120° F. K_i	Assume t = 120° F. x_i	Assume t = 140° F. K_i	Assume t = 140° F. x_i	Assume t = 131° F. K_i	Assume t = 131° F. x_i
N_2	3.20	67.0	0.05	78.0	0.04	73.0	0.04
CH_4	57.75	11.5	5.0	13.5	4.3	12.6	4.6
C_2H_6	16.47	1.63	10.1	1.85	8.9	1.75	9.4
C_3H_8	11.49	.675	17.0	.78	14.7	.74	15.5
C_4H_{10}	6.55	.304	21.6	.373	17.6	.34	19.3
C_5H_{12}	2.73	.132	20.7	.175	15.6	.155	17.6
C_6H_{14}	1.26	.055	22.9	.076	16.6	.066	19.1
C_7H_{16}+	0.55	.031	17.8	.044	12.5	.038	14.5
Total	100.00	..... Temp. too low	115.15	..... Temp. too high	90.2	..... Temp. correct	100.0

Bubble-Point Calculations

The problem in bubble-point calculations is the same as that for dew-point calculations, except that now the known quantities are the x_i instead of the y_i. So in this case one again takes L = O.

In using the first nomograph the proper reference point on the reference line is found as before but this time it is connected with the given x value and the vapor composition y read on the x_c scale on the left.

In using the second nomograph, one starts from the given x value on either the upper or lower scale, whichever is more convenient, and moves vertically downward or upward until he reaches the horizontal line which meets the curve for the assumed K value at the L = O line. The resulting point of intersection is on a x_c line which represents the vapor composition.

In either case the process is repeated until a temperature and pressure at which the condition

$$\sum_{i=1}^{n} y_i = 100 \tag{6}$$

holds is determined.

Advantages Cited

As mentioned before, each nomograph has its advantages. The first nomograph, with its expanded K scales, is more accurate, especially for K values above unity. Also, being an alignment chart, interpolation on it is generally easier since the eye can more readily interpolate between two points on a linear scale than between labeled contour lines as employed on the second nomograph.

On the other hand, for quick approximations to the solution, the second nomograph is useful because it is not necessary to differentiate between the various K scales and because one can see at a glance approximately what effect a change in the value of L or K will have on the resulting x value.

Vaporization of Petroleum Fractions

A Method of Calculation for the Vapor-Liquid Equilibria of Complex Mixtures

W. D. HARBERT
Lion Oil Company, El Dorado, Ark.

THE calculation of liquid-vapor equilibria plays an important part in the design and operation of process plants and equipment.

In general, the calculation of pressure-temperature-composition relations and distillation conditions involves two types of mixtures. The first of these is the component mixture for which the composition is commonly expressed as the percentage of the compounds present. Thus a natural gas is said to contain so much methane, ethane, etc. The other type is the complex mixture, which contains a large number of compounds and for which the composition is commonly expressed as a batch distillation curve. Thus a kerosine is said to boil between 350 and 550° F. Such refinery charge stocks and products as crude oils, gasolines, kerosines, naphthas, gasoils, and residuals are usually described thus in terms of their boiling ranges.

A method using simple calculus is presented for the calculation of vapor-liquid equilibria for complex mixtures. The compilations are based on the true boiling point curve of the mixture and are performed with a plot of dM/dT_B vs. T_B. The procedures and equations used are similar to those for component mixtures. The results obtained can be made exact within the accuracy of the data. Four illustrative examples are given of calculations involving a single equilibrium.

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IN GENERAL the calculation of pressure-temperature-composition relations and distillation conditions involves two types of mixtures. The first of these is the component mixture, for which the composition is commonly expressed as the percentages of the compounds present. Thus a natural gas is said to contain so much methane, ethane, etc. The other type is the complex mixture, which consists of a large number of compounds and for which the composition is commonly expressed by a batch distillation curve. Thus a kerosine is said to boil between 350° and 550° F.

For component mixtures the use of either Raoult's law or vapor-liquid equilibrium constants, K values, is common for the calculation of such vaporization properties as the dew-point temperature.[5, 6, 10, 13, 16] There is a complete system of equations and procedures for these computations, and the results can be made as accurate as the data used. The basic equations used for these single equilibrium calculations can be written as follows:

Vapor-Liquid Equilibrium

$$y = Kx \tag{1}$$

or

$$Vy = (VK/L)\,(Lx) \tag{2}$$

Over-all Material Balance

$$F = V + L \tag{3}$$

Component Material Balance

$$Fz = Vy + Lx \tag{4}$$

Compositions Summations

$$\Sigma Fz = F \qquad \Sigma Vy = V \qquad \Sigma Lx = L \tag{5}$$

A mathematical system for complex mixtures comparable with the one for component mixtures would seem to be desirable, particularly for refinery process

NOMENCLATURE

A	value of Lx for general component in still liquid for example IV and Equation 9
B	value of Lx for reference component (component $K = 1.0$) in still liquid for Example IV and Equation 9
D	liquid density, pounds per gallon
F	total moles of original mixture
Fz	total original amount of a component present (dM/dT_B for complex mixtures)
K	vapor-liquid equilibrium constant, defined by equation $(Vy) = (VK/L)\,(Lx)$ for complex mixtures
L	total moles of liquid in the equilibrium
Lx	amount of a component present as liquid (dM/dT_B for complex mixtures)
M	number of moles
mw	molecular weight
T	temperature of equilibrium, ° F.
T_B	boiling point temperature, ° F., obtained from TBP for complex mixtures
V	total moles of vapor in the equilibrium
Vy	amount of a component present as vapor (dM/dT_B for complex mixtures)
v	liquid volume percent of total mixture
x	mole fraction of component in liquid
y	mole fraction of component in vapor
z	mole fraction of component in total original mixture

Subscripts

a	assumed value, on L/V
c	calculated value, on L/V
1	conditions at beginning of distillation in Equation 9
2	conditions at any time during distillation in Equation 9

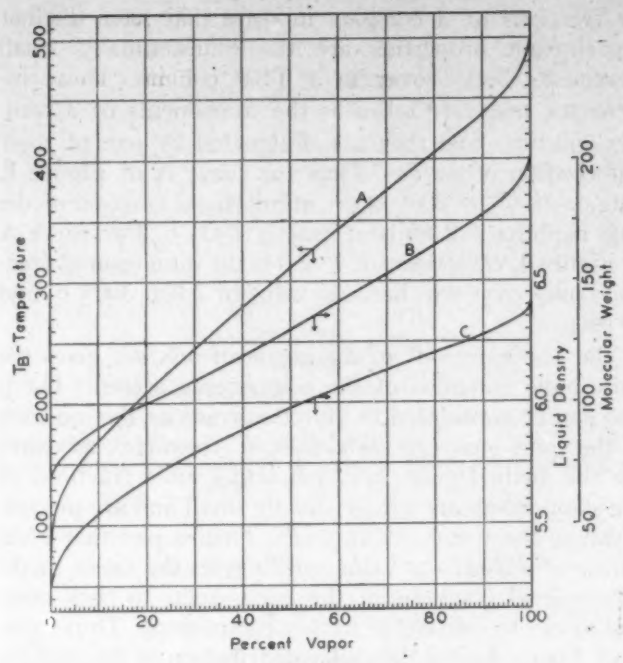

FIGURE 1

Properties of a Petroleum Naphtha.

A = true boiling point curve; B = molecular weight of fractions distilled over; C = liquid density, pounds per gallon, of fractions distilled over.

work. In general, this type of mixture is either considered to be divided into artificial compounds having a short boiling range[3, 6, 8, 9, 17, 18] or treated empirically by methods based on large amounts of laboratory data.[4, 11, 12, 14] The use of differential curves for complex mixtures has been frequently proposed,[3, 6, 7, 17] but the actual details required to make these calculations practical are not commonly known. The present paper gives the simple calculus for one method by which complex mixtures may be handled by equations and procedures similar to those used for component mixtures. The results obtained with this method can be made exact within the accuracy of the data used.

Plot of dM/dT_B vs. T_B

For component mixtures the fundamental vaporization data are commonly given in terms of the components present. Thus it is said that the K value of *n*-hexane is 4.5 at 300° F. and 20 pounds per square inch absolute.[3] But for a complex mixture the number of compounds is large, so that some basis for the vaporization data other than molecular species is necessary. A common and well developed laboratory determination for complex mixtures is the true boiling point curve, or TBP, which is obtained by the batch distillation of the mixture in a multiplate column at a high reflux ratio. This curve is usually given as liquid volume percent distilled over plotted against column top temperature or boiling point (v against T_B). For the method of this article the vaporization data of a complex mixture will be based on the boiling points as given by a TBP. Thus it will be said that the K value of the component that boils at 156° F. (*n*-hexane) is 4.5 at 300° F. and 20 pounds per square inch absolute. Values of K based on boiling temperature have been published by White and Brown.[18]

But a TBP as such cannot be used in the vapor-liquid equilibrium calculations; consequently it must be changed into a form that can. For direct use in the computations by the method of this article, the TBP is put into the form of the differential curve dM/dT_B *vs.* T_B, where M represents moles and T_B is the boiling temperature. This curve provides the connections between total moles, component moles, and vaporization properties that are required for the calculations. The simple calculus of obtaining this curve from a TBP will be given, and after this the properties and uses of the curve will be described and illustrated.

The basic relation used in obtaining a value of dM/dT_B is the identity

$$\frac{dM}{dT_B} = \frac{dv}{dT_B} \times \frac{dM}{dv} = \frac{dv}{dT_B} \times \frac{D}{mw} \qquad (6)$$

where the values of the derivatives are for the infinitesimal fractions of the mixture boiled over at a temperature in a TBP column. Values of dM/dT_B are plotted at the corresponding values of T_B to give the desired curve. As an example a plot of dM/dT_B against T_B will be obtained for the petroleum naphtha of Figure 1. Curve A is the true boiling point curve for this mixture and gives the boiling temperatures of the liquid fractions distilled over. Curves B and C give the molecular weights and the densities of these liquid fractions, respectively. The steps in obtaining a plot of dM/dT_B against T_B from these data are as follows:

Step 1. Obtain a plot of dv/dT_B for the mixture. The slopes of the TBP, curve A of Figure 1, are measured, and the reciprocals, dv/dT_B, are plotted against T_B. The resultant points are carefully averaged and smoothed. The final values are given by curve A of Figure 2. Thus at 200° F. the measured slope of the TBP is 3.66, and the reciprocal of this, $dv/dT_B = 0.273$, is plotted at $T_B = 200°$. As a convenient check the area under this curve represents liquid volume and should total 100.

Step 2. Obtain the plot of dM/dT_B against T_B with Equation 6. Values of dv/dT_B, mw, and D, all at the same value of T_B, are substituted in Equation 6 to give a value of dM/dT_B. Points of dM/T_B against T_B obtained in this way are plotted and smoothed to give curve B of Figure 2. As an example at 200° F., dv/dT_B is 0.273, and, from curves B and C of Figure 1, $mw = 98$ and $D = 5.67$. Then

$$100\,\frac{dM}{dT_B} = 100 \times 0.273 \times (5.67/98) = 1.58 \qquad (6)$$

and this value is plotted on curve B of Figure 2 at $T_B = 200°$ F.

A principle in the solution of vapor-liquid equilibrium problems is that the entire calculation can be based on dimensionless ratios such as L/V and K. Any absolute values can be made to cancel out in the calculations. Thus in the compositions reported for component mixtures it is only necessary that the ratios of moles of compounds to each other be known; the actual numbers of moles are not needed. For the same reason it is permissible to change the decimal point on the values of dM/dT_B from 0.01 to 1.0, and this is done in the remainder of this article. Similarly Equation 6 does not need to be dimensionally correct. The actual scale used for the values of dM/dT_B plotted against T_B is unimportant so long as the shape of the curve is not affected. Other short cuts based on this use of ratios will be given. These may not seem to be desirable mathematically but they save time and are the way the calculations are actually performed.

It may be desired to do the opposite of the operation just described—that is, to obtain a TBP from a plot of dM/dT_B against T_B. This is done by a reversal of the steps given. Values of dM/dT_B and of D and mw are substituted in Equation 6 to obtain values of dv/dT_B, and these are plotted against T_B as in curve A of Figure 2. The area under this curve represents v, liquid volume. The graphical integration of successive sections of the curve with a planimeter gives the TBP.

In order to use a plot of dM/dT_B in the calculations it is first necessary to know the K values at the conditions

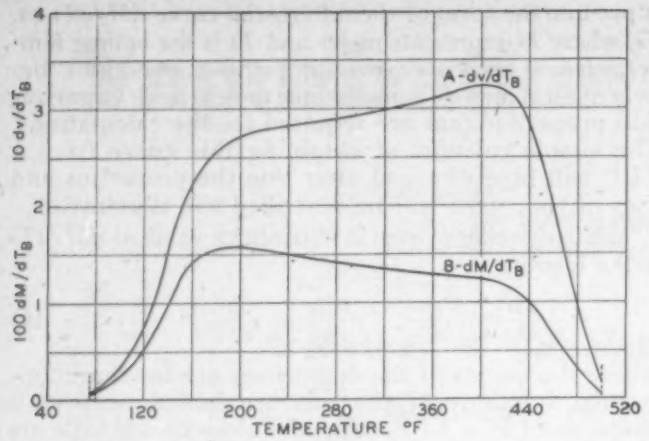

FIGURE 2

Differential Curves for Petroleum Naphtha Plotted Against Boiling Temperature.

A = dv/dT$_B$, reciprocal slope of TBP, vs. T$_B$ from curve A of Figure 1. B = dM/dT$_B$ vs. T$_B$ as obtained from Curve A with Equation 6.

of the equilibrium as a function of the boiling temperature. This relation for the conditions of 300° F. and 20 psia is given by curve A, Figure 3, which is a straight line plot of T_B against K on semilogarithmic paper. This curve was obtained by plotting boiling temperature against the known K values of 300° F. and 20 pounds for the hydrocarbons of lower molecular weight,[3] and then averaging the resultant points as a straight line extending over the desired temperature range of 75° to 500° F. In using a plot of dM/dT_B it is desirable to be able to read the K values directly from the figure. A convenient way of doing this is to draw vertical lines at values of T_B and label them with the corresponding vales of K. A construction of this type is illustrated in Figure 4, where curve A represents the original naphtha and is a copy of curve B, Figure 2. The K values at the boiling temperatures were read from curve A, Figure 3.

A plot of dM/dT_B against T_B marked off with K lines, such as Figure 4, was chosen for actual use in the complex mixture calculations because of certain very useful properties. First, the area under a plot of dM/dT_B *vs.* T_B represents moles and can be easily and quickly determined with a planimeter. Second, the shape of this curve depends on the mixture and is independent of the conditions of the equilibrium. For new conditions of temperature and pressure new sets of vertical K lines are drawn. Third, the K values for the calculations are read directly from the figure, and the results obtained are plotted on the corresponding K lines. The use of these properties will be illustrated in the example calculations to be given.

Analogy with Component Mixture Calculations

A complete mathematical system for the calculation of the vapor-liquid equilibria for complex mixtures could be given independently of the already existing system for component mixtures. However, the calculations are quite similar in many respects, and it is believed that it would be helpful to develop the analogy between them. In order to do this it is necessary that the terms used for component mixtures be defined for complex mixtures in such a way that both can be spoken of with the same words and calculated by the same equations.

For component mixtures the "components" are simply the compounds present, and each has a single distinct set of vaporization properties. For the method given here the fractions of a complex mixture that have distinct vaporization properties are the infinitesimally small increments boiled over in a TBP column. These increments, then, are taken as the components of a complex mixture, and they are designated by one of their vaporization properties. Thus for curve A of Figure 1, component $T_B = 200°$ is the infinitesimal fraction of the total naphtha that boils at exactly 200° F. For curve A of Figure 4, component $K = 2.0$ is the infinitesimal fraction boiled over that has a K value of 2.0 at 300° F. and 20 psia.

The "composition" of a component mixture gives the finite mole fractions of the components present; but it also can be considered to give the ratios of the amounts of the components to each other. For complex mixtures for the method given here the actual mole fractions of the components are infinitesimally small and are proportional to the values of dM/dT_B. Then a plot that gives values of dM/dT_B at values of T_B gives the ratios of the infinitesimal fractions of the components to each other and so can be considered to be a composition. Thus curve A of Figure 4 gives the molar distribution of the naphtha of Figure 1 along the boiling range and so can be considered to represent the composition of the naphtha.

For component mixtures the terms Fz, Lx, and Vy represent moles of the components and are used to calculate divisions of the total component between vapor and liquid. For complex mixtures, as already stated, the amounts of the components are infinitesimally small and are proportional to values of dM/dT_B. But these values of dM/dT_B can also be used to obtain the division of the components between vapor and liquid. Then for complex mixtures the terms Fz, Lx, and Vy are taken as the corresponding values of dM/dT_B as read from a composition curve. As an example, for the total original naphtha of curve A, Figure 4, the value of dM/dT_B for component $K = 4.0$ is 1.44, and this can be stated simply as $Fz = 1.44$. This definition permits Equations 2 and 4 to be used for complex mixtures exactly as for component mixtures, and this is illustrated in the example calculations to be given later. The terms x, y and z are not used alone for complex mixtures.

For component mixtures the total mole terms—F, L, and V—are obtained by the summation of the moles of the individual components; but a complex mixture is considered to consist of an infinite number of components, so that an integration rather than a summation is required to obtain total moles. This integral, which corresponds to Equation 5, can be written as

$$\int \left(\frac{dM}{dT_B}\right) dT_B = M \qquad (7)$$

where M represents moles and can be either F, L, or V. In actual use this integral is the area under a composition curve (dM/dT_B *vs.* T_B) and can be easily and quickly determined with a planimeter. As already discussed, it is the ratios of the total mole terms to each other and not their absolute values that are necessary for the calculations. Then in determining moles as areas, no integrating machine factor need be used, provided it is the same for all of the composition curves, as it will cancel out in the calculations. Then as an example, the planimeter reading for the area under curve A, Figure 4, is 18.67, and this can be stated simply as $F = 18.67$.

In summary, then, the infinitesimal components of complex mixtures can be divided between vapor and liquid and summed to obtain total moles exactly as are

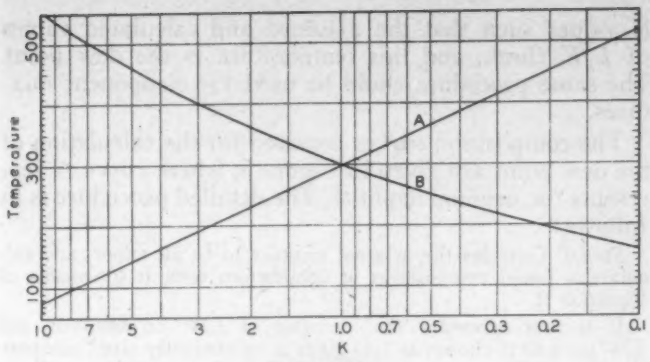

FIGURE 3

Vaporization Data for Petroleum Naphtha.

A = boiling temperature (T_B) vs. K at 300° F. and 20 pounds per square inch absolute (straight line on semilogarithmic paper); B = temperature of vapor-liquid equilibrium (T) vs. K at 20 pounds for component for which K = 1.0 at 300° F.

the finite components. The same equations and procedures can be used for both types of mixtures.

Constant Relative Volatility

A useful assumption that is frequently made for vapor-liquid equilibrium calculations for component mixtures is that of constant relative volitility.[16] In this it is considered that the ratios of the K values remain constant over a temperature range, however much the K values themselves change. With this assumption the temperature unknown is either eliminated completely from the calculation or its solution is obtained without trial and error. This gives a quicker, more direct calculation. With the use of some care in averaging the K value ratios, the errors due to this assumption can be kept small for regular systems of chemically similar compounds such as the saturated hydrocarbons.

This assumption of constant K value ratios is also useful in the complex mixture calculations. With it the K values at a temperature can be obtained by multiplying the known K values at another temperature by a constant. In order to do this it is first necessary to know the K value-*vs.*-temperature relation for one component. Component $K = 1.0$ is chosen as this reference component for the naphtha composition of curve A, Figure 4, and the desired plot of K values against the temperature of the equilibrium for this component is given by curve B, Figure 3. This curve was obtained by plotting the reciprocal of the absolute temperature against the K value at 20 psia on semilogarithmic paper for compounds for which the K values are known, and then drawing a straight line of the same slope through the point at which $T = 300°$ F. and $K = 1.0$. Values of T *vs.* K from this curve are then replotted on semilogarithmic to give curve B, Figure 3. As an example of the use of this curve, the K value of component $K = 4.0$ of curve A, Figure 4, will be found at 260° F. and 20 pounds. From curve B, Figure 3, for reference component $K = 1.0$, the K value at 260° F. is 0.58. Then with the assumption of constant K value ratios, the K value of component $K = 4.0$ at 260° F. is (4.0/1.0) (0.58) = 2.32. The K values of other components and at other temperatures can be obtained similarly.

The method given here is not dependent on any of the assumptions made for the vaporization data. It is only necessary that the K values be known as a function of the boiling temperature, and any source of these data including direct laboratory determinations can be used. As with component mixtures, the assumption of constant relative volatility is a convenient way of obtaining K values at different conditions, but it is not essential to the method.

Example I. Equilibrium Division Into Vapor and Liquid

The first calculation is to determine the division of the naphtha of Figure 1 between vapor and liquid at equilibrium for the conditions of 295° F. and 20 psia. The principal equation used is

$$(Lx) = \frac{(Fz)}{1 + (VK/L)} \qquad (8)$$

which is obtained by combining Equations 2 and 4. The procedure used is to assume a value of L/V for the equilibrium, and, with the Equation 8 and the known K values, to calculate a division of the naphtha into vapor and liquid. From this a new calculated value of L/V is obtained. This step is repeated until the assumed and calculated values of L/V check. Exactly the same procedure is used for component mixtures.

The composition curves required in the calculation are given in Figure 4, and curve A represents the original naphtha. The detailed procedure is as follows:

Step 1. Assume a reasonable value of L/V, and, with Equation 8 and the K values, calculate a division of the naphtha into vapor and liquid.

The value of L/V assumed for the first trial is 1.0. At 295° F. the K value of reference component $K = 1.0$ is found to be 0.93 (curve B, Figure 3), and the K values of the other components change in this same ratio of 0.93 to 1.0 (constant relative volatility). Then for component $K = 3.0$, the K value at 295° F. is $3.0 \times 0.93 = 2.79$. From curve A, Figure 4, the value of dM/dT_B at $K = 3.0$ is $Fz = 1.57$, and Equation 8 becomes

$$Lx = \frac{1.57}{1 + (2.79/1.0)} = 0.415 \qquad (8)$$

This value of Lx is plotted at $K = 3.0$ to give one point of curve B, Figure 4. The other points at the K values shown were obtained similarly. This curve gives the required division of the original naphtha into liquid and vapor for an assumed L/V of 1.0.

Step 2. Obtain a new, calculated L/V from the curve of step 1. The moles of liquid present are represented by the area under curve B, Figure 4, and the value is given by the planimeter as $L = 9.07$. The corresponding moles of total mixture is the area under curve A, Figure 4, and has been found to be 18.67. Then by Equation 3, $V = F - L = 18.67 - 9.07 = 9.60$, and the new calculated L/V is $9.09/9.60 = 0.946$. Evidently this is not sufficiently close to the assumed value of $L/V = 1.0$, and a new trial must be made.

Step 3. Repeat steps 1 and 2 until the assumed and calculated *values of* L/V check.

The third and final assumption was $L/V = 0.72$, and the result-

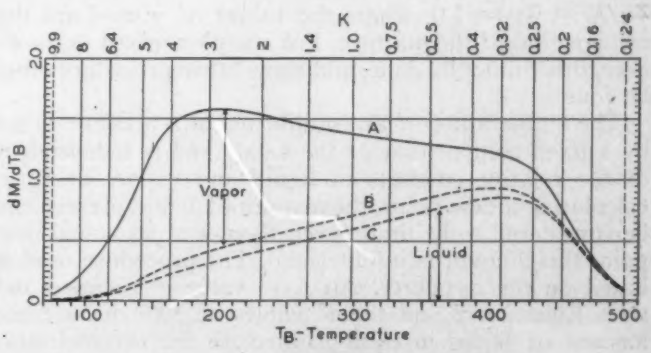

FIGURE 4

Composition Curves for Example I.

A = composition curve for original naphtha; B = division of naphtha into vapor and liquid for first trial; C = division of naphtha into vapor and liquid for final trial.

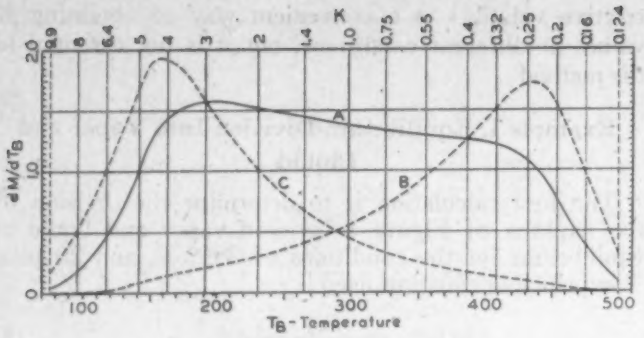

FIGURE 5

Composition Curves for Exaple II.

A = composition curve for original naphtha; B = composition of liquid in equilibrium with original naphtha as vapor (dew point); C = composition of vapor in equilibrium with original naphtha as liquid (bubble point).

ant division of the naphtha is given by curve C, Figure 4. The moles of liquid present are represented by the area under this curve, and the value is found to be $L = 7.84$. Then $V = F - L = 18.67 - 7.84 = 10.83$, and the new calculated L/V is $7.84/10.83 = 0.724$. This is taken as sufficiently close to the assumed value of $L/V = 0.72$, so that no further trials are required. Then curve C, Figure 4, can be considered to represent the final division of the original naphtha into vapor and liquid at 295° F., and 20 psia.

Step 4. Convert the mole fractions obtained in step 3 into liquid volume percent.

For this a plot of dv/dT_B *vs.* T_B is prepared for the values of curve C, Figure 4, by means of Equation 6. The area under this curve represents liquid volume and is given as $v = 15.15$ by the planimeter. The corresponding area under curve A, Figure 2, for the total original mixture is $v = 32.79$. Then the liquid naphtha of curve C, Figure 4, is $100 \times (15.15/32.79) = 46.2$ liquid volume percent of the total.

Then as a final answer the naphtha of Figure 1 is $100 - 46.2 = 53.8$ percent vaporized at 295° F. and 20 pounds. This is a point on the equilibrium flash curve for the mixture (curve B, Figure 9). The TBP values of the liquid and vapor of the equilibrium are given by curves B and D of Figure 6, respectively. These were obtained from curves A and C of Figure 4 by integrating plots of dv/dT_B *vs.* T_B as already described.

Example II. Dew and Bubble Point Temperatures

The next values to be calculated for the naphtha of Figure 1 are the dew and bubble point temperatures at 20 psia. The dew point is the temperature at which the first drops of liquid condense in the cooling of the total mixture as vapor, and the bubble point is the temperature at which the first bubbles of vapor form in the heating of the total mixture as liquid. In solving for the dew point of a component mixture, the usual procedure is to find the temperature that gives K values such that $\Sigma y/K = \Sigma x = 1.0$, where the values of y used are the composition of the mixture. For complex mixtures, however, this cannot be done, and some other procedure must be found.

The temperature of an equilibrium at a pressure is set by a fixed composition of the vapor and is independent of the relative amounts of liquid and vapor. Then, in calculating a dew point, the amount of liquid present can be considered to be finite, even though at an actual dew point this amount is infinitesimal. The procedure used is based on this property. An L/V value is assumed and with Equation 2 and the K values at 300° F., a finite amount of liquid in equilibrium with the original mixture as vapor is calculated. With the assumption of constant K value ratios the composition of this liquid is independent of the temperature chosen. From the amount of this liquid, calculated values of L and L/V are obtained. The temperature is then found which gives K values such that the assumed and calculated values of L/V check, and this temperature is the dew point. The same procedure could be used for component mixtures.

The composition curves required for the calculation of the dew point are given in Figure 5, where curve A represents the original naphtha. The detailed procedure is as follows:

Step 1. Consider the original mixture to be all vapor, and calculate a liquid composition in equilibrium with it by means of Equation 2.

It is first necessary that a value of L/V be assumed, and $L/V = 0.40$ is chosen as this gives a conveniently sized composition curve. For component $K = 0.55$ the value of $Vy = Fz = 1.29$ is read from curve A, Figure 5, and Equation 2 becomes

$$Lx = 1.29\ (0.40/0.55) = 0.94 \qquad (2)$$

This value of Lx and others similarly obtained are plotted to give curve B, Figure 5, which is the required composition of the liquid in equilibrium with the original mixture as vapor.

Step 2. Determine the new calculated values of L and L/V. The moles of the original mixture as vapor are represented by the area under curve A, Figure 5, and the value has been found to be $V = F = 18.67$. The corresponding moles of liquid are shown by the area under curve B, Figure 5, and the value is given by the planimeter as $L = 10.90$. Then the new, calculated value of L/V is $10.90/18.67 = 0.584$, as compared with the assumed value of 0.40 used in step 1.

Step 3. Find the temperature (dew point) that gives K values such that the assumed and calculated L/V values would check.

If the K values at 300° F. used in Equation 2 in step 1 were multiplied by the ratio $(L/V)_c/(L/V)_a$ and the calculation repeated, then the assumed and calculated values of L/V would be the same, and the K values would be at the dew point temperature (constant K value ratios). Then for reference component $K = 1.0$ the value of K at the dew point temperature is

$$K \times (L/V)_c/(L/V)_a = 1.0\ (0.584/0.40) = 1.46$$

From the plot of K *vs.* T for this component (curve B, Figure 3) the temperature corresponding to $K = 1.46$ is 333° F., and this is the dew point temperature for the naphtha at 20 psia.

The solution for the bubble point temperature is similar, except that liquid and vapor are interchanged. The original mixture is considered to be all liquid and a

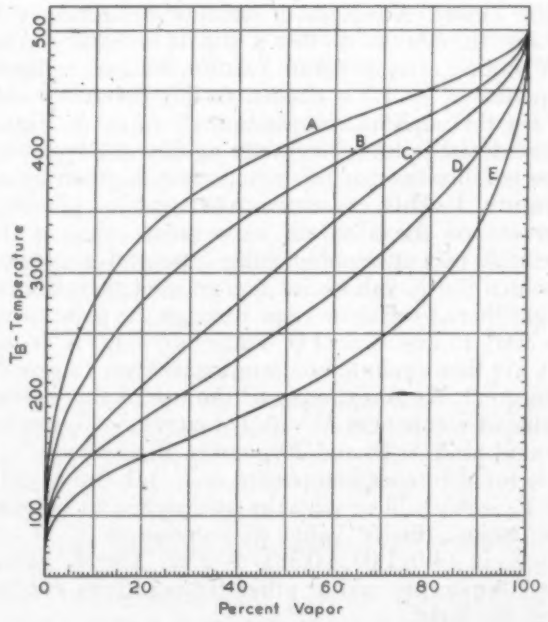

FIGURE 6

True Boiling Point Curves.

A = TPB of liquid in equilibrium with original naphtha as vapor; B = TPB of portion of naphtha that is liquid at 295° F. and 20 pounds per square inch; C = TPB of original naphtha; D = TPB of portion of naphtha that is vapor at 295° F. and 20 pounds per square inch; E = TPB of vapor in equilibrium with original naphtha in liquid.

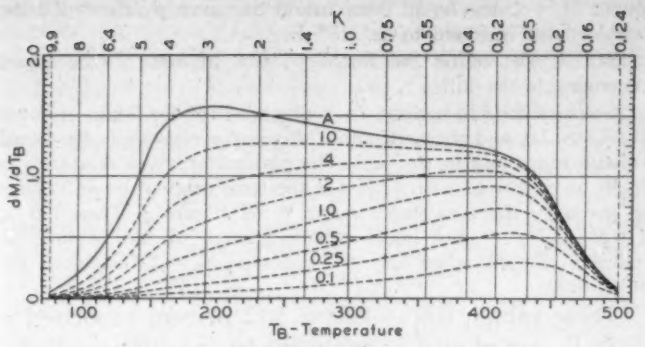

FIGURE 7

Composition Curves for Example III.

A = composition curve for original naphtha; broken lines give divisions between vapor and liquid and are labeled with value of L/V assumed in calculation.

vapor in equilibrium with it is calculated by means of Equation 2. L V is assumed to be 3.0, and the resulting vapor composition is given by curve C, Figure 5. The moles of vapor are $V = 11.42$, the calculated value of L/V is $18.67/11.42 = 1.63$, and the ratio $(L/V)_c/(L/V)_a$ is $1.63/3.0 = 0.54$. Then the K value of reference component $K = 1.0$ is $1.0 \times 0.54 = 0.54$ at the bubble point, and from curve B, Figure 3, this temperature is 254° F.

These two temperatures, 254° and 333° F., are shown as the end points of the equilibrium flash curve for the mixture (curve B, Figure 9). The calculated TBP of the liquid in equilibrium with the original naphtha as vapor is curve A, Figure 6, and the corresponding TBP for the vapor in equilibrium with the original naphtha as liquid is curve E, Figure 6. These were obtained from curves B and C of Figure 5 by the method already described.

Example III. Equilibrium Flash Curve

An important curve in the study of complex mixtures is the equilibrium flash curve, which gives the liquid volume percent of the mixture that is vaporized against the temperature of the system (T *vs.* v) for a constant pressure. The present calculation is to obtain the equilibrium flash curve at 20 pounds psia for the naphtha of Figure 1. One point on this curve was obtained in example I—namely, at 295° F. and 20 pounds the naphtha is 53.8 percent vaporized. The calculation of a point on the equilibrium flash curve is quicker than the calculation of example II, as no specific temperature need be set, and consequently the trial and error is eliminated. The genral procedure used to find a point on this curve is as follows:

A value of L/V and a temperature are assumed, and an equilibrium division of the total naphtha into vapor and liquid is obtained with Equation 8. The temperature required for this equilibrium (constant K value ratios) and the percent of the total naphtha that is vapor are then determined. These values are the coordinates of one point on the desired curve. Exactly the same procedure would be used for component mixtures.

The composition curves required for the calculation are given in Figure 7, where curve A represents the original naphtha. The other curves give divisions of the naphtha between vapor and liquid and are labeled with the values of L/V used to obtain them. The detailed calculation for an assumed value of $L/V = 0.5$ is as follows:

Step 1. Assume a value of L/V, and, with Equation 8 and the K values at 300° F., calculate an equilibrium division of the naphtha into vapor and liquid.

For this particular case L/V is taken as 0.5. For component $K = 3.0$ the value $Fx = 1.57$ is read from curve A, Figure 7, and Equation 8 becomes

$$Lx = \frac{1.57}{1 + (3.0/0.5)} = 0.224 \qquad (8)$$

This value is plotted on the $K = 3.0$ line to give one point of curve $L/V = 0.5$. The other points of the curve and the other curves of this figure were obtained similarly. These curves give equilibrium divisions of the total mixture into vapor and liquid.

Step 2. Determine the temperature required for the equilibrium of step 1.

If the K values at 300° F. used in step 1 were multiplied by the ratio $(L/V)_c/(L/V)_a$ and the calculation repeated, then the assumed and calculated values of L/V would check, and the K values would be at the temperature of the equilibrium (constant K value ratios). The planimeter gives the area under curve $L/V = 0.5$ as $L = 6.32$, and the corresponding area under curve A of Figure 7 for the total original naphtha is $F = 18.67$. By Equation 3, $V = F - L = 18.67 - 6.32 = 12.35$, and the new calculated L/V is $6.32/12.35 = 0.51$ as compared with the assumed value of $L/V = 0.50$. Then for reference component $K = 1.0$, the K value at the temperature of the equilibrium is $K \times (L/V)_c/(L/V)_a = 1.0 \times (0.51/0.50) = 1.02$. From curve B, Figure 3, the temperature corresponding to this value of K is 301° F., and this is the temperature of the equilibrium.

Step 3. Determine the liquid volume per cent vaporized in the equilibrium of step 1.

For this a curve of dv/dT_B *vs.* T_B is prepared for the liquid of curve $L/V = 0.5$ by means of Equation 6. The area under this curve represents liquid volume and is found to be $v = 12.41$. The corresponding area under curve A of Figure 2 for the total mixture is $v = 32.79$. Then the mixture is $100 \times (12.41/32.79) = 37.9$ liquid volume percent liquid and 62.1 percent vapor for the conditions of the equilibrium.

Then the final values obtained for an assumed L/V of 0.5 is 62.1 percent vaporized at 301° F., and this point is plotted on the equilibrium flash curve of the mixture (curve B, Figure 9). The other points of this curve were obtained similarly and are given in Table 1. The end temperatures, the bubble and dew points, are from Example II. These points are nearly a straight line. The original naphtha of Figure 1 is a comparatively sharp cut, so that the tails frequently present on flash curves do not occur to an appreciable extent.

Example IV. Differential Distillation Curve

As an example of a more complicated calculation the differential distillation curve for the naphtha of Figure 1 will be obtained. In a differential distillation the mixture is boiled, and the vapors, which are in equilibrium with the remaining liquid, are removed as fast as formed. The final curve is plotted as boiling temperature against

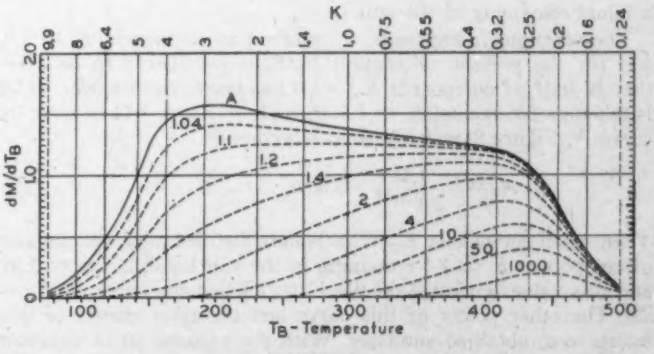

FIGURE 8

Composition Curves for Example IV.

A = composition curve for original naphtha; broken lines give liquids remaining in the still and are labeled with value of B_1/B_2 assumed in calculation.

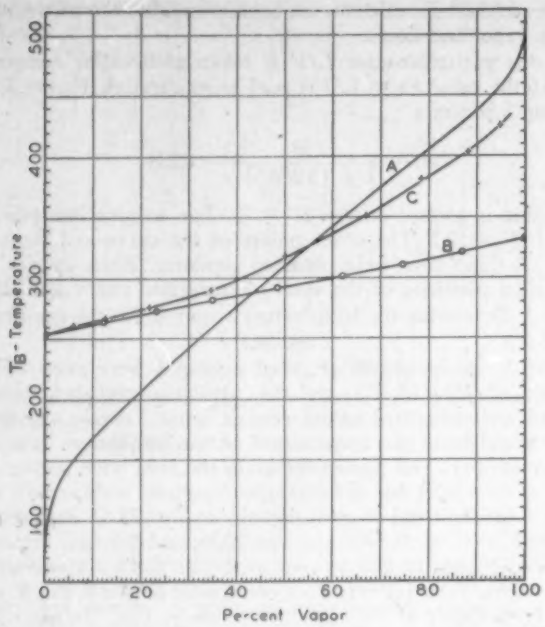

FIGURE 9

Distillation Curves for Petroleum Naphtha.

A = TPB of naphtha; B = equilibrium flash curve from Examples II and III; C = differential distillation curve from Examples II and IV.

liquid volume percent distilled over (T *vs.* v). The basic equation used is the integrated form of the Rayleigh equation:[15]

$$(A_1/A_2) = (B_1/B_2)^{K_A/K_B} \qquad (9)$$

where A is any component and B is the reference component. The ratios A_1/A_2 and B_1/B_2 are the reciprocals of the fractions of these components that remain in the still. In order to obtain one point on the differential distillation curve, a value is assumed for B_1/B_2, and the resultant composition of the liquid remaining in the still is found with Equation 9. The boiling temperature and the liquid volume percent of this still liquid are then determined. These two values give the coordinates of one point on the differential distillation curve. The same procedure would be used for component mixtures.

The composition curves for this calculation are given in Figure 8, where curve A represents the original naphtha of Figure 1.

The detailed calculation for an assumed value of B_1/B_2 *of* 2.0 is as follows:

Step 1. Assume a value of B_1/B_2 and, with Equation 9, calculate a liquid remaining in the still.

The reference component, B, is taken as component $K = 1.0$, and for the present calculation B_1/B_2 is considered to be 2.0—that is, half of component $K = 1.0$ has been distilled over. If A is taken as component $K = 1.4$, then $A_1 = Fz = 1.44$ is given by curve A, Figure 8, and Equation 9 becomes

$$A_2 = \frac{1.44}{2.0^{1.4/1.0}} = 0.56 \qquad (9)$$

Then when component $K = 1.0$ is half distilled over, the amount of component $K = 1.4$ remaining in the still liquid is $Lx = 0.56$, and this value is plotted on the $K = 1.4$ line for curve $B_1/B_2 = 2.0$. The other points of this curve and the other curves of this figure were obtained similarly. With the assumption of constant relative volatility the ratio K_A/K_B is independent of the still temperature.

Step 2. Determine the boiling temperature (bubble point) of the liquid composition obtained in step 1.

The details of this calculation have already been given in example II. For the liquid composition of curve $B_1/B_2 = 2.0$ the bubble point is found to be 318° F.

Step 3. Determine the liquid volume percent of the liquid remaining in the still.

A plot of dv/dT_B against T_B is prepared for the liquid of curve $B_1/B_2 = 2.0$, and the area under this curve represents the liquid volume remaining in the still. The planimeter gives this area as 16.36, as compared with 32.79 for the total original liquid volume as given by the area under curve A of Figure 2. Then $100 \times (16.36/32.79) = 49.9$ liquid volume percent of the mixture remains in the still when half of component $K = 1.0$ has been distilled over.

These values, $100 - 49.9 = 50.1$ percent vaporized at 318° F., are plotted as one point on the differential distillation curve for the naphtha curve C of Figure 9. The other points shown for this curve were obtained similarly. The initial point is the bubble point of the original mixture as obtained from Example II. The numerical values obtained for the points are listed in Table 2.

A laboratory distillation that roughly approximates differential distillation is the ASTM[1] or Engler, in which the mixture is placed in a flask and boiled off. There is a basic similarity in shape between a typical ASTM curve and curve C of Figure 9. The low initial temperatures of an ASTM curve are due partially to the air present in the system at the beginning of the distillation.

Discussion

To a person familiar with the details the methods given here are both direct and quick. They require less time than the corresponding discontinuous methods, in which the complex mixture is divided into artificial components having short boiling ranges. Thus a point on a composition curve can be found as easily as a mole fraction and averaged temperature can be obtained for artificial component. In determining total moles, a planimeter can obtain an area almost as quickly as a column of figures can be added. Values of Lx and Vy can be as easily set down by a point on a curve as by a number in a table. The dew point calculation of Example II can be completed in less than ten minutes of comparatively slow work, with seventeen points used to determine the liquid

TABLE 1

Equilibrium Flash Curve (Curve B, Figure 9)

Assumed L/V	v, Liquid Vol. Percent as Vapor	T, Temp. of Equil., ° F.
∞ (bubble point)	0	254
10	12.3	264
4	23.1	273
2	34.8	282
1.0	48.5	292
0.5	62.1	301
0.25	74.6	311
0.10	86.8	321
0 (dew point)	100.0	333

TABLE 2

Differential Distillation Curve (Curve C, Figure 9)

Assumed B_1/B_2	v, Liquid Vol. Percent Distilled Over	T, Still Temp., ° F.
1.0 (bubble point)	0	254
1.04	5.9	261
1.1	12.5	267
1.2	21.7	276
1.4	34.6	290
2.0	50.1	318
4	66.9	351
10	78.4	381
50	88.0	403
1000	94.8	425

composition in equilibrium with the original mixture as vapor.

Another advantage of the method is that the differential plots of the compositions give a much clearer picture of an equilibrium or distillation than do the corresponding columns of figures of the discontinuous methods. Thus a comparison of Figures 7 and 8 shows that a differential distillation gives a better separation of the naphtha into its high and low boiling components than the corresponding equilibrium flash. The greater slopes of the lines dividing vapor and liquid in Figure 8, as compared with Figure 7, mean that less heavy constituents are removed in the vapor, and less light constituents remain in the liquid for a differential distillation. An additional advantage of the methods given here is that the use of point values and the exact calculus is more desirable theoretically than is the corresponding use of averaged values and arithmetic in the discontinuous methods. If more and smaller finite divisions are used in the discontinuous calculations, the errors are reduced, but, also, the time required is increased.

Pure compounds can easily be combined with complex mixtures in these calculations. A convenient way of doing this is to plot the compound as a vertical line at the correct boiling temperature on a composition curve such as curve A, Figure 4. Length on this line can be considered to represent moles and so is the equivalent of so much area (moles) under the composition curve of the complex mixture.

In the differentiation of a TBP to obtain a plot of dM/dT_B *vs.* T_B certain minor difficulties arise. The TBP should be drawn as a smooth curve that averages the points of the laboratory data, and any plateaus should be either smoothed out or removed and treated as single compounds. The individual values of dv/dT_B plotted against T_B should also be averaged as a smooth curve. The values of dv/dT_B at the low and high boiling ends of the TBP are difficult to measure directly. These portions of the curve are best handled by extending the plot of the measurable values of dv/dT_B in a smooth curve that approaches the T_B axis as an asymptote. Any vapor loss should be added to the low boiling and any residue to the high boiling end of the curve. With reasonable care errors due to these and other causes can be made negligible.

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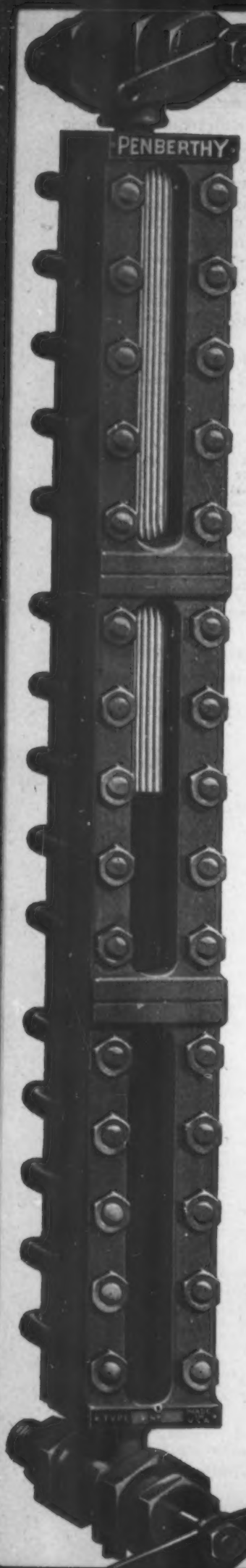

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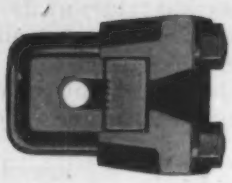

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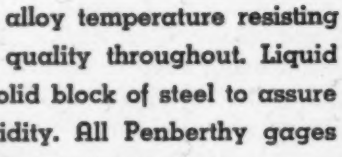

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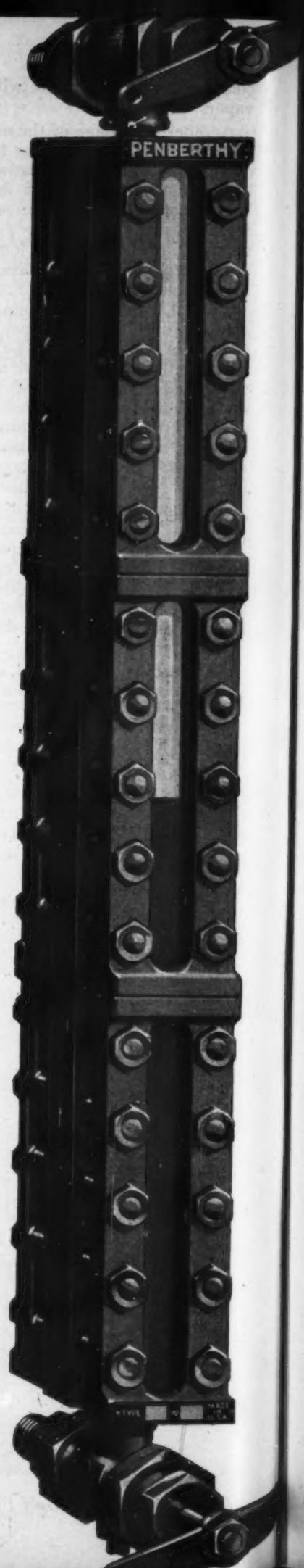

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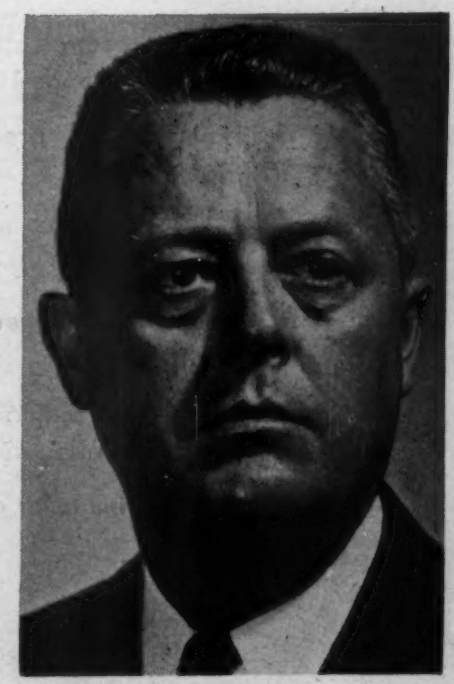
Trio of Texaco Men Contribute Paper

THREE Texas Company men, John T. Wiley, Joe E. Deloney and S. Walter Denton, all connected with the refining plant at Port Arthur, Texas, worked out the paper "Accurate Particle-Size Determination of Fluid Catalyst by Coordination of Roller Analysis and Microscopic Examination" which starts on page 90 of this issue. This was one of the presentations at a Division of Refining Symposium on Cracking Catalysts before the annual meeting of the American Petroleum Institute last month in Chicago.

John T. Wiley, born in Mobitee, Texas, in 1907, attended public schools and West Texas State College in Canyon, Texas, graduating in 1928 with a Bachelor of Science degree in chemistry. He worked for Independent Engineering Company at Borger, Texas, from 1928 to 1929 as chemist and in plant operations. From 1929 to 1932 he was plant chemist and chief chemist for MacMillan Refining Company at Borger, and in 1932 he went to Texhoma National Gas Company as chemist. Here he remained until going with The Texas Company at Port Arthur as analytical chemist in 1936. In 1942 he was made assistant chemist-in-charge of the analytical laboratory and under his guidance a number of procedures for analyzing fluid catalysts have been developed.

Joe E. Deloney was born in 1911 in Athens, La., where he attended the public schools. He was enrolled at Centenary College, Shreveport, La., for two years but obtained his B.A. degree in science from Louisiana Tech, Ruston, La. Starting with The Texas Company in 1935, he worked on the experimental cracking units in the laboratory until 1938, at which time he came to the analytical laboratory as chemist where he has helped develop a number of analytical methods, some of which pertain to the analysis of fluid catalysts.

S. Walter Denton was born in Trinidad, Texas, in 1919, where he attended the public schools. He received his B.S. in chemistry from Trinity University, Waxahachie, Texas, and his M.S. from Texas Tech, Lubbock. Upon completing school he obtained a position with The Texas Company in 1942 and since that time has worked as an analytical chemist in the laboratory at Port Arthur. Much of his time has been spent on the development of analytical methods in connection with fluid catalysts.



E. Q. CAMP

Math Teacher Busy At Sour Crude Job

E. Q. CAMP, author of "Sour Crudes —Equipment Protection" was educated in Texas schools (1928 graduate of Sam Houston State Teachers College, Huntsville) and taught chemistry and mathematics before joining Humble Oil & Refining Company 18 years ago. There, almost from the start, he found that he had traded classroom troubles for something equally as involved inasmuch as he has devoted his major efforts with Humble to the solution of corrosion problems brought about particularly through handling of sour crudes from certain Texas oil fields.

At present Camp, who is senior research chemist for Humble, is giving his full time to the direction of research work involving corrosion and refinery waste disposal problems. His paper, which starts on page 100, made a good impression when presented at Chicago last month before a Division of Refining group of the American Petroleum Institute.

Nomographs the Work Of Two Phillips Men

THE "Nomographs for Vapor-Liquid Equilibrium," a most valuable article which starts on page 128 of this issue, is the work of two Phillips Petroleum Company Research Department men, Fred H. Poettmann and R. B. Rice.

Poettmann joined Phillips' research

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4. High exhaust pressure trip actuating emergency trip valve.
5. Remote control electrical trip actuating emergency trip valve.
6. Throttling type emergency trip valve for hand throttling.
7. Gland seal piping for condensing operation—prevents air leakage into turbine casing.
8. Gland sealing for operation with gas.

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DISTRICT OFFICES IN PRINCIPAL CITIES

H-718

staff in January, 1946. He received his undergraduate training in chemical engineering at Case Institute of Technology, graduating with the B.S. degree in 1942. He obtained his Sc.D. in chemical engineering in 1946 at the University of Michigan.

Since joining Phillips, Poettmann has been engaged in research on problems connected with reservoir engineering. He is secretary-treasurer of the Oklahoma Section of the American Institute of Chemical Engineers and is a member of the ACS and AIME.

Rice joined Phillips in the summer of 1941 as reservoir engineer in the production department. From 1941 to 1945 he was employed as assistant in the department of mathematics, Ohio State University, but in 1945 he returned to Phillips as a process and design engineer in the gasoline department. Since January, 1946, he has served as a mathematician in the geophysical section of the Research Department. He is a member of the Mathematical Association of America, American Mathematical Society and Society of Exploration Geophysicists.

FRED H. POETTMANN

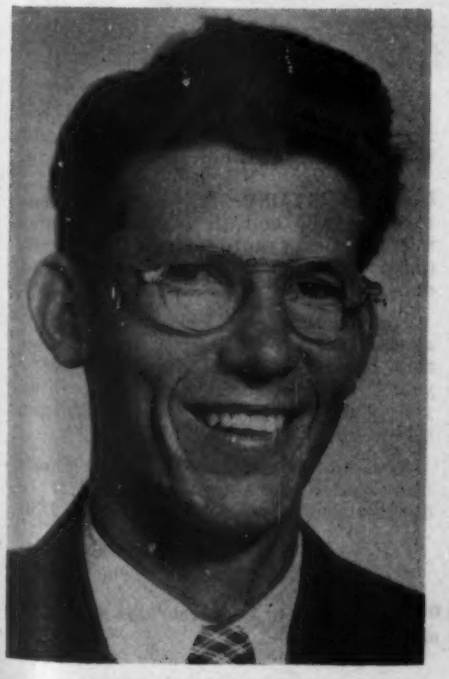
R. B. RICE

Texas Teachers Contribute Article

A FIVE-YEAR-OLD daughter and a new son in July are among the avocations listed by John Griswold, co-author with Warren B. Brooks of "High Range Vapor Pressures and Antoine Equation Constants of Some C_3 and C_4 Hydrocarbons" (starts on page 122). His other off-the-job pursuits include gardening, swimming, camping and meteorology.

Griswold is professor of chemical engineering at the University of Texas, Austin, where he has been since 1936. His research in problems of distillation, fractionation and vapor-liquid equilibria has been particularly along lines as applied to recovery of petroleum hydrocarbons. In these fields he has to his credit some 30 technical publications and is the author of the textbook "Fuels, Combustion and Furnaces." He is regarded as a technical expert and witness on gas explosions and fires. Director of research, RFC Office of Rubber Reserve research project at the University of Texas. Griswold holds a B.S. degree in chemical engineering from Illinois (1928) and Doctor of Science in Chemical Engineering, Massachusetts Institute of Technology, 1932. He was connected with The Pure Oil Company as a chemist before taking up teaching.

Brooks, the other author, holds a B.S. in chemical engineering from Oklahoma A. & M., (1944), an M.S. in chemical engineering from the University of Texas (1947) and currently is working on his Ph.D. as he holds these positions at Texas: tutor in chemistry, instructor in industrial management and research assistant, Rubber Reserve Research project. His hobbies did include a playing interest in football until he suffered a broken nose while practicing for an intramural game at Austin. Now, he says, he prefers picnicking, popular and classical music and study of time and motion. Maybe—who knows?—he got that broken nose because he had failed to devote sufficient hours study to "time and motion."



WARREN B. BROOKS

JOHN GRISWOLD

Lion Oil Engineer Is Oklahoma Grad

THE lead article in this month's PETROLEUM REFINER (page 85) deals with the Thermofor catalytic cracking unit which recently was put in operation by Lion Oil Company at El Dorado, Ark. The description and discussion is by F. L. Emert, plant engineer.

Emert is a graduate of the University of Oklahoma, located at Norman, where he received his degree in 1936 in refinery engineering. From the university, he went on to Bartlesville, Okla., where he joined Phillips Petroleum Company. With this company he did general refinery engineering at Bartlesville, transferring later to Borger, Texas, where plant operation engaged his attention.

Emert joined Lion Oil Company, his present employer, in the spring of 1940. General engineering, plant layout and equipment selection have kept him busy here, along with process design and refinery engineering.

As to hobbies, Emert admits "sports in general," but when pinned down shifted right easily into an elaboration regarding hunting and fishing. So the streams around El Dorado see much of him in season and now that he has bought himself a new 5-horsepower outboard motor, we wouldn't be surprised if they didn't see him even more frequently.

At home Emert has a young son Frankie, who manages, despite refinery and field and stream duties, to claim a part of papa's time.

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PACKLESS CORRUGATED EXPANSION JOINTS

Challenge to Crude Oil

U. S. producers must effect economies if impending competition of oils from natural gas and coal is met

L. J. LOGAN

JUST as the U. S. through wartime necessity made itself virtually independent of natural rubber by manufacturing a substitute "raw rubber" from materials at hand, so the country could if necessary make itself independent of natural crude petroleum within a few years by synthetically making substitute "petroleum" from any of several available materials.

It is not only scientifically possible to substitute natural gas or coal for crude oil in making liquid fuels, but natural gas already is actively competitive with crude as a raw material for the petroleum industry, while coal quite possibly may become a commercially employed source of liquid fuels within five to ten years, being already in the pilot plant stage of such utilization.

Several commercial scale plants for obtaining gasoline and other liquid fuels from dry natural gas already are under construction or planned, including that at Brownsville, Texas, and one scheduled for the Hugoton gas field, Kansas. Proved U. S. reserves of natural gas are equivalent in heat content to those of crude oil.

Also under construction, near Pittsburgh, is a pilot plant for converting coal into gas, with the ultimate aim of then converting such "synthesis gas" into gasoline and other liquid fuels by the same process as that to be used in the Texas and Kansas natural gas-to-gasoline plants. U. S. reserves of coal are so great as to make the proved reserves of crude oil and natural gas appear almost insignificant by comparison. Coal represents 98.8 percent of the country's mineral fuel energy reserves, excluding fissionable materials; oil shale constitutes 0.8 percent; petroleum 0.2 percent; and natural gas 0.2 percent.

While oil shale reserves are relatively large, and while oil is being obtained from shale in some petroleum-deficient countries, the production of oil from shale so far has proved difficult and costly and it will be resorted to in the U. S. only if much better methods are developed.

Self-Sufficiency Assured

The present potential independence of the U. S. from natural crude petroleum, which heretofore was a vital essential, is a fact of tremendous significance to the nation and all its citizens, because it gives assurance that the U. S. A. need never be conquered or commercially exploited for lack of liquid fuels. While crude petroleum reserves of the U. S. and of the rest of the world are of continuing importance, they are not a matter of life or death for the nation or even of continued health and prosperity. If our crude reserves look too small and difficult to maintain in the face of enormous present day and prospective requirements, and if we refuse to lean too heavily upon foreign crude reserves, it is necessary only to reduce our reliance upon crude of any origin and to begin depending at least partially upon coal. The problem of reserves would thereby be solved rather permanently, as U. S. coal reserves are equivalent to the nation's fuel needs for more than 1000 years.

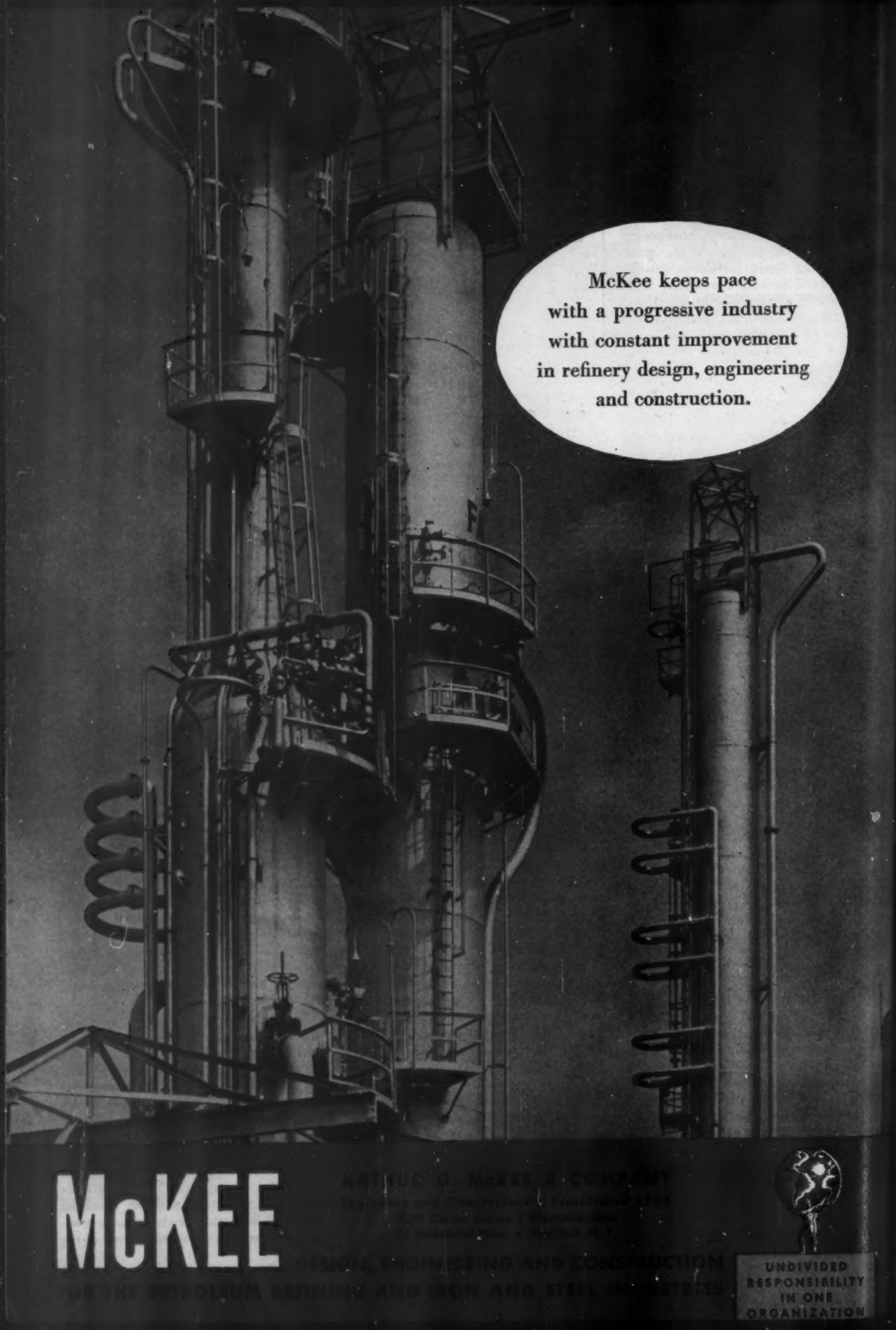
Instead of bothering to drill for crude at excessive depths, in remote and inaccessible places, upon the storm-swept open sea of the Continental Shelf, or under other prohibitively difficult conditions, the petroleum industry fortunately has an alternative. If the less accessible potential crude reserves are too difficult or costly to develop, the industry can leave them undeveloped and begin using as its raw material, in larger and larger proportion, first natural gas and eventually coal.

So important and so promising is the future role of synthetically produced liquid fuels that encouragement of the development of such fuels is almost unanimously agreed upon as a vital point in the national policy on petroleum. Both large and small oil companies as well as key government agencies and officials consider it wise to develop synthetic oils to supplement natural petroleum as promptly as possible, although many Americans very much dislike the thought of encouraging or even permitting large-scale imports of crude to supplement domestic, believing that a strong and self-sufficient domestic fuels industry is a vital necessity in national defense.

So thoroughly is the importance of synthetic oils accepted within the petroleum industry itself that some of the larger oil companies, which are spending millions of dollars annually in research, are devoting around one third of all such expenditures to research in synthetic fuels. While they thereby recognize the crude supply as an increasingly important problem, they also thereby assert their determination not to let the problem defeat them—the determination to keep on meeting any demand for liquid fuels that may develop, by going to substitute raw materials if necessary.

While the development of synthetic liquid fuels serves the interest of the nation and the industry by affording the possibility of reducing dependence on crude petroleum, this development also has a secondary value in that it will force a maximum of progressiveness and efficiency in the continuing production of natural petroleum. By threatening to supplant oil from natural petroleum, the new oil from natural gas and oil from coal will make it imperative for the producers of natural petroleum and their engineers and scientists to work out the most efficient and cheapest methods of drilling and production. For the engineers of crude producing companies, synthetic fuels issue a challenge that can be met only by further advances in drilling and producing techniques. As the synthetic processes are further developed upon commercial basis, they may be expected to make the synthetic fuels available at lower costs, while it will be difficult to prevent the costs of oils from natural petroleum from further increasing, in view of the need for drilling deeper and other manifestations of greater difficulty in proving up new crude reserves. In prospective competition between producers of synthetic and natural oils, the crude oil producers thus are at a disadvantage that can be offset only by improvements in existing methods and practices.

While competition between synthetic oils and natural petroleum is only in its initial stages, the synthetic oils from dry natural gas by the Fischer-Tropsch process must be recognized as already actively competitive with oils from crude petroleum. The first two projected plants, in Texas and Kansas, will be full-blown commercial "gas refineries," costing in excess of $15 million each and producing from natural gas about 7500 barrels per day of high octane gasoline, alcohols, diesel oil, and other products, with gasoline representing about 80 percent of the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total and distillate fuel oil 13 percent. The plants will commercially apply the same methods used for three years by Standard Oil Company (N. J.) in a pilot plant at Baton Rouge, La., making 20 to 30 barrels per day of liquid fuels.

When the Texas plant was announced about two years ago (prior to decontrol of prices) the owners estimated that gasoline would be produced at a cost of about 5½ cents a gallon, about the same as the cost then involved for gasoline refined from crude. While the cost perhaps should be figured higher now, because of increased plant construction costs, it is possible that capital costs may be held down or even reduced as experience in operating the plants points the way to improvements in plant design. Natural gas as the raw material is only about 20 percent of the cost of the gasoline produced, as against 50 percent for capital charges, indicating that gas prices would have to increase substantially before appreciably raising the cost of gasoline from gas. Even if the total cost of gasoline from natural gas is estimated to have increased 25 percent above the 5½ cents of two years ago, in line with the increase meanwhile in construction costs, then a present cost of only 7 cents a gallon is indicated.

In contrast, crude prices in the past two years increased 90 cents a barrel or nearly 2¼ cents a gallon, adding that much to the crude oil refiner's cost of raw material alone, in addition to other increases in his costs. The average price of crude in the U. S. having reached $2.13 per barrel, unrefined crude itself now costs 5 cents a gallon at the well, before transportation to the refinery. Because of higher crude and operating costs, Mid-Continent refiners of gasoline from crude have had to increase their quotations to a range of 8 to 10¼ cents a gallon (November 15, 1947), depending on octane rating. Apparently, the cost of gasoline from crude oil now appreciably exceeds the cost of gasoline from natural gas, making gas-to-gasoline plants quite attractive.

200,000 Barrels by 1956

The initial 15,000 barrels per day of synthetic oil production from natural gas is expected before 1950, and such production has been estimated by a qualified oil company economist at 200,000 barrels daily by 1956. On this basis, from the standpoint of meeting demand for gasoline, the future "gas refineries," using natural gas as the raw material, would wipe out demand for crude oil by 30,000 barrels daily in 1950 and by 400,000 barrels daily in 1956, allowing for a 40 percent gasoline yield from crude as at present. In the "gas refinery" the dry natural gas is first broken down by partial burning into carbon monoxide gas and hydrogen gas. Those gases then are fed into a synthesis reactor, where they combine, in the presence of a finely powdered catalyst, to form the liquid fuels.

In view of the projected production of oils from natural gas, it may be stated as a general rule that henceforth it will be profitable to drill a contemplated oil well only if oils made from the crude oil it yields involve overall costs no greater than those of oils derived from natural gas. It remains to be seen, for example, whether or not cheap and plentiful natural gas as a source of oils will discourage or make impractical any efforts to produce natural oils at depths of 15,000 or 20,000 feet or from structures beneath the open ocean along the Continental Shelf.

Oil from coal in the U. S. will be merely a variation of the Fischer-Tropsch process, using coal instead of natural gas as the source for the synthetic oils. Instead of partially burning natural gas into carbon monoxide and hydrogen and then catalytically converting those gases into liquid fuels, the oil-from-coal refinery will mine and finely crush large tonnage of coal and convert the fine coal, with aid of steam and oxygen or air, into carbon monoxide and hydrogen gases, which then are converted catalytically into the liquid fuels. To get the carbon monoxide and hydrogen from coal instead of natural gas obviously is the harder way because of requiring the mining and crushing operations for several thousands of tons of coal per day in the case of a large commercial plant instead of connecting up with a natural gas pipe line.

Although oil from coal is only in the pilot plant stage and not yet competitive with oil from crude petroleum, the coal and oil interests which are now actively developing the process envisage the first commercial-scale plant production of oil from coal by 1950-1951. The actively competitive stage for oil from coal may be no more than five or ten years away, although it has not been definitely set down on the calendar by some forecasters until 1970.

The pilot plant near Pittsburgh for gasifying coal, now under construction, will cost about $300,000, process 50 tons of coal daily, and produce about 2,400,000 cubic feet of synthesis gas per day. That synthesis gas could be then converted into about 120 barrels per day of synthetic gasoline, but that extra step will not be taken at the pilot plant, for it would be essentially the same as in the conversion of natural gas into gasoline, while a pilot plant at Linden, N. J., already is experimentally producing 15 gallons per day of liquid fuels from synthesis gas.

"Coal Refinery"

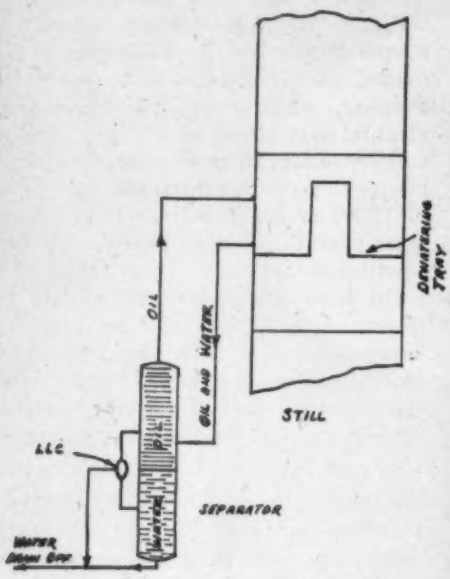
After the pilot plant near Pittsburgh shows the way to gasify coal, there will be built the first commercial-scale "coal refinery." It is envisaged as a mammoth plant, costing perhaps $120 million and doing a complete job of supplying gaseous or liquid fuels, taking in coal from the mouth of the mine and turning out either high-value gas or liquid fuels for convenient marketing in the adjacent region. While this synthesis gas will not be low enough in cost to compete with natural gas from the Southwest, it is expected to command premium prices for meeting emergency or extraordinary winter demands, and at other seasons the bulk of the output will be converted into liquid fuels.

So far, oil from natural petroleum has some competitive advantage over oil from coal. It has been estimated by one oil company research man that crude oil could rise another $1 per barrel in price before running into serious competition from coal as the raw material for liquid fuels. Another major oil company has estimated that motor fuel could be produced from coal for between 7 and 10 cents per gallon, while still another authoritative study put the cost at 6½ cents a gallon, or at a level definitely competitive with motor fuel produced from crude oil.

Any competitive advantage that crude oil now holds over coal is not great enough to justify complacency among crude oil producers or petroleum engineers as to the present methods of drilling and the prevailing methods of operating oil fields. The costs of producing oil from coal are virtually certain to come down as the methods now contemplated are perfected or as new and better methods are developed. Offering the possibility of great savings in making oils from coal, for example, is the underground gasification of coal, which has been tried experimentally with some success and which would cut out the heavy costs of mining and crushing the coal. Meanwhile, as brought out previously, the costs of producing oils from natural petroleum threaten to further increase rather than to come down because it has become necessary to drill deeper, drill in less accessible places, resort to secondary recovery, and assume other extra costs in the increasingly difficult effort to find and develop as much crude oil as the nation is consuming, with usage at unprecedented proportions and constantly expanding.

OPERATION HOW TO DO IT MAINTENANCE

The drawoff and return lines at dewatering tray level.



Separator for removing water from stripping still.

The new way--the separator placed at grade level.

HOW TO—

Make Efficient a Plant Water Draw-Off System

STRIPPING stills of gasoline plants, where steam is used to strip light fractions from absorption oil, often require a dewatering tray to remove the condensed steam from the fractionating section of the still. The top tray of the still may be maintained at a temperature lower than that required to maintain vaporization conditions for steam. The steam condenses and necessarily must be removed from the system.

Some of the heavier fractions of the overhead product will also condense at the dewatering tray temperature, and will be drawn off with the water. A separator is necessary to recover the oil and return it to the tower. General practice is to place the separator in an elevated position just below the tray level alongside the tower. This may require scaffolding for installation, a permanent platform for control operation, and even temporary scaffolding for some maintenance operations.

By placing the separator at grade level, Shamrock Oil and Gas Corporation, at its McKee plant gasoline unit located at Sunray, Texas, has somewhat lowered the cost of initial installation and has greatly increased the convenience of operation for the water drawoff system.

The above sketch shows the flow detail of the dewatering tray and separator while the photographs picture the actual installation.

HOW TO—

Transport Foam-Type Fire Extinguishers

A HANDY fire extinguisher carrying case similar to a sample basket as shown in the accompanying photo offers a solution in the matter of transporting foam-type extinguishers. The case prevents the extinguishers from turning over while in transit in a truck and two men are able quickly and easily to carry six extinguishers at a time. Any convenient material can be used to construct the case and only a small amount of welding time is required. Adequate extinguisher facilities thus can be available at all times.

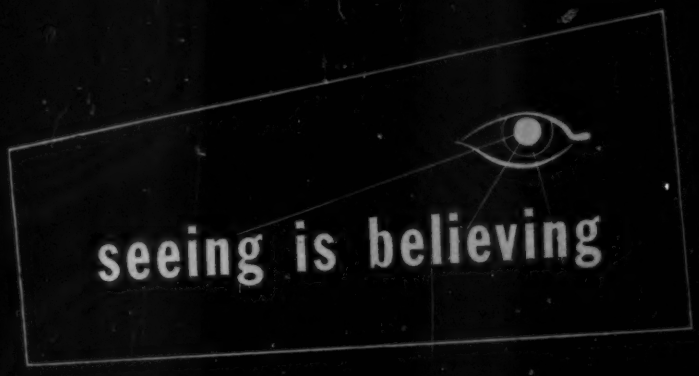
Handy device for carrying fire extinguishers.

HOW TO—

Insulate Boiler Fronts Permanently

A METHOD of insulating boiler fronts that eliminates the necessity of tearing out and reinsulating them each time inspection or maintenance is required, is being used by Shamrock Oil & Gas Company at its Sunray, Texas, plant.

Blanket type insulation incased in lightweight sheet metal is fabricated into sections to fit around the piping detail of the boiler fronts. The sections are about one inch in thickness and any standard wool type of insulation will adequately serve the purpose. Bolt clips



The chart shows the progressive acceptance of Petro-Chem Iso-Flow Furnaces . . . first introduced in 1940, today more than 425 are in service and more significantly they are operating in all phases of petroleum and petro-chemical processing.

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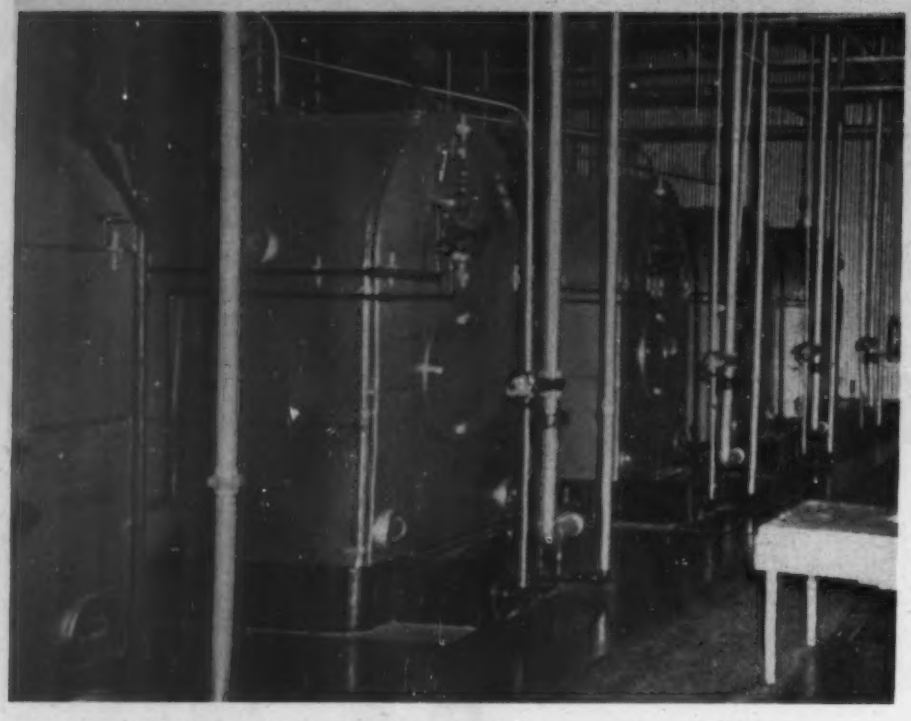


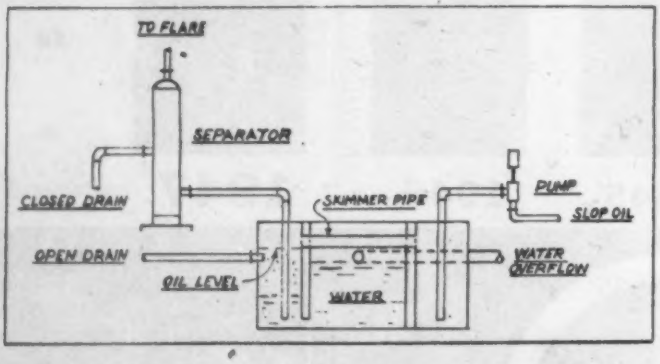
Permanent boil insulation affords easy access for inspection and maintenance

spaced at convenient intervals hold the sections in place. Unbolted, the sections are easily removed, may be stacked out of the way during the work period, and are easily and quickly replaced on the boilers.

These sections require little or no maintenance and quickly pay dividends on the initial cost as they last indefinitely with normal care.

HOW TO—

Construct Gasoline for Gasoline Plant



Schematic design for construction of plant skimmer basin.

VARIOUS types of skimmer basins are employed in gasoline and cycling plants to provide the necessary separation of oil and water from the drainage systems. In all cases, one of the most important prerequisites of the systems is their ability to operate with a minimum of maintenance and attention from plant personnel.

Separation in First Basin

The skimmer basin setup illustrated in the accompanying sketch was constructed to meet this prerequisite. Discharge lines from the dual drainage system of the plant, closed and open drains, enter the basin. The closed drain system, which is under pressure, discharges into a separator. Gas from the separator is flared and the oil and water flow by gravity through a line to the bottom of the skimming basin. Material from the open drain enters the basin below the oil level. The oil and water separate in the first section of the basin, and the oil rises to the top and flows through the skimmer pipe, which serves as the weir, into the pump suction part of the basin to be pumped to slop oil tanks. Water below the oil flows under the separating wall into the center section of the basin and is removed through the overflow line to sewage disposal.

HOW TO—

Cut Dismounting Time for Bearings

A MORE economical and time-saving method of removing large anti-friction bearings from machines—in minutes instead of hours—is being applied by an increasing number of manufacturers, according to Dr. Haakon Styri, director of research for SKF Industries, Inc.

By pressuring oil from a hand-operated hydraulic pump between the bore of the bearing and the shaft on which it is mounted, the inner ring expands and "floats" on the oil film, enabling the operator to remove the bearing with greater ease and in less time. The method is also successful for coupling two shafts together.

The application of oil at 7000 to 8000 pounds per square inch reduces dismounting time to 10 or 15 minutes, as compared with several hours required by the former method involving use of pullers and hydraulic jacks.

HOW TO—

Protect Laboratory Distillation Thermometers

MANY laboratories use counterbalanced pulley assemblies attached to their distillation thermometers to help prevent the breakage of the thermometers. By building a rack at a convenient height above the laboratory distillation bench and using the thermometer shipping tube as a shield, further protection is provided. Both ends of the tube are removed so the pulley lines extend through the tubes. When the thermometers are not in use they are placed in the tubes completely protected. Labels on the tubes identify the thermometer range.

Rack protects thermometers in laboratory

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Record Attendance at Annual Meeting of American Petroleum Institute at Chicago

Each company and individual in the petroleum industry will have to share actively in the responsibility of promoting good will for the industry if its public relations program is to succeed fully, President William R. Boyd, Jr., stated in opening the general session of the 27th annual meeting of the American Petroleum Institute, held last month in Chicago.

Another important responsibility resting upon each member is that of promoting harmony within the industry and tolerance of its members toward other members, he declared.

"Oil men have too much work to do to fritter away their energy in family squabbling that ordinarily does little more than give encouragement to those who would mess up our industry." Boyd asserted, "If we could only find some way to debunk a lot of trivial things and dephlogisticate anyone who stirs up trouble within the industry, it seems to me that our customers, our public, would have a much better opinion of our industry."

Especially is there danger, said Boyd, in seeking government intervention in controversial affairs within the industry because this is the way toward restrictive and stifling governmental controls.

Regarding the difficult situation of the present supply shortage, Boyd definitely refused any suggestion of government action. Rationing instituted by the government would not increase the supply to the public and by the same token would cause the public to lose confidence in the petroleum industry, he thought. "The industry is more able to do a better job—the present situation is not due to an oil shortage but to an excess in demand at a time when industry as a whole is handicapped by the lack of materials," he pointed out.

In an address which replaced one originally scheduled by Secretary of the Interior J. A. Krug, who was absent because of illness, Under-Secretary Oscar L. Chapman stated that the public welfare demands that the petroleum industry and government agencies dealing with oil matters work together even more fully than heretofore. The present situation of the oil industry would be far worse than it is, Chapman pointed out, if the Interior Department's Oil and Gas Division had not collaborated on the tanker problem last year.

Large scale exports of petroleum products as contemplated under the Marshall Plan will call for real sacrifice on the part of the American people, Walter S. Hallanan, chairman of the National Pe-

Meetings

December

12—Natural Gasoline Association of America, regional meeting, Amarillo, Herring Hotel.

12-13—American Chemical Society, Southwest Region, Houston.

29-30—Fourteenth Annual Chemical Symposium, American Chemical Society, Chicago.

January, 1948

12-16—Society of Automotive Engineers, annual meeting, Detroit, Book-Cadillac Hotel.

February

16-19—American Society of Chemical Engineers, regional meeting, New Orleans.

March

24-26—Natural Gasoline Association of America, annual convention, Ft. Worth, Texas Hotel.

April

5- 7—Western Petroleum Refiners Association, annual meeting, Galveston, Galvez and Buccaneer Hotels.

5- 8—National Association of Corrosion Engineers, annual conference-exhibit, St. Louis, Jefferson Hotel.

19-21—American Society of Lubrication Engineers, annual convention, Buffalo, Statler Hotel.

21-23—National Petroleum Association, Cleveland, Hotel Cleveland.

28-30—Liquefied Petroleum Gas Association, annual convention, Kansas City.

May

4- 5—American Gas Association, Natural Gas Department, Houston, Rice Hotel.

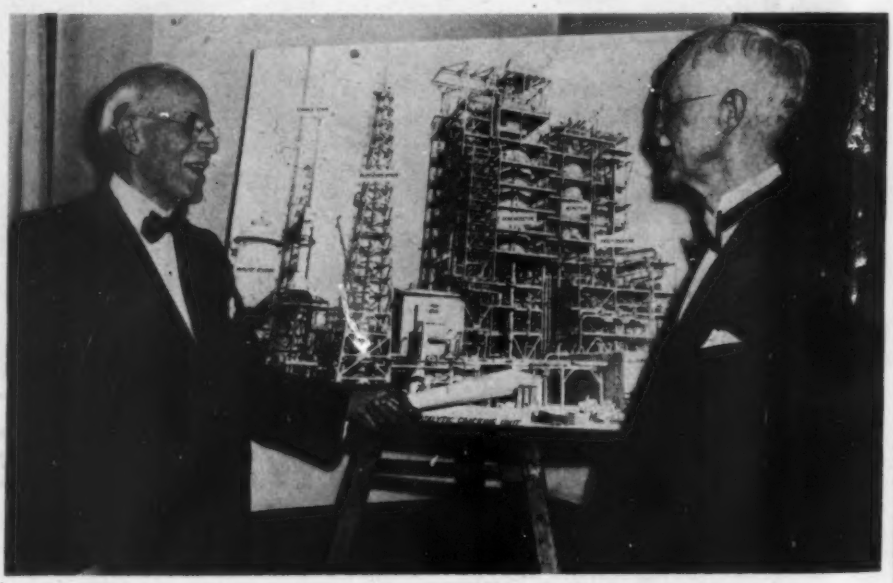
16-22—International Petroleum Exposition, Tulsa.

24-26—American Gas Association, joint production and chemical conference, Asbury Park, N. J., Berkeley-Carteret Hotel.

June

21-25—American Society for Testing Materials, annual meeting, Detroit, Book-Cadillac Hotel.

Burton Given API Medal of Achievement



Dr. William M. Burton, left and Dr. Robert E. Humphreys pictured when they were honored a year ago as "pioneers in oil cracking" by the American branch of the Newcomen Society of England at a dinner in the University Club of Chicago. In the background is a photograph of one of the latest developments in petroleum technology, the fluid catalytic cracking unit which went on stream that month at the Whiting, Ind., refinery of Standard Oil Company (Indiana). Dr. Burton last month was presented with the American Petroleum Institute Gold Medal for Distinguished Achievement and a citation which read in part: "You were a pioneer in demonstrating the value of laboratory research and testing to assure delivery of satisfactory petroleum products in the market place. You saw—almost 40 years ago—the need for major changes in refining methods to meet the stupendous demand for gasoline that the automobile was to create. You developed the first commercially successful cracking process, more than doubling the potential yield of gasoline available from crude oil. This process marked the industry's advance from merely separating and purifying the natural ingredients of petroleum to altering its chemical composition so as to produce tailor-made molecules for specific purposes. . . . Modern American, living and moving on wheels and shafts lubricated and largely powered by petroleum, owes honor to you for your scientific foresight and for qualities of character and intelligence that showed the way to a new era for petroleum. . . ." Dr. Burton received his A.B. degree from Western Reserve in 1886 and his Ph.D. degree from Johns Hopkins in 1889. In 1890 he entered the employ of the Standard Oil Company (Indiana) as its first chemist at the Whiting refinery. He became general superintendent in 1895, a member of the board of directors in 1911, and served as vice president from 1915 until 1918, when he became president. He resigned the presidency of the company in 1927, but continued in an advisory capacity as a member of the board of directors until 1928. With Dr. Humphreys as a colleague, Dr. Burton started the first work ever done in an industrial laboratory toward producing petroleum products by chemical change rather than physical separation. Dr. Burton has received many high honors, among them the Willard Gibbs medal of the American Chemical Society and the Perkin medal of the Society of Chemical Industry. The latter is the ultimate recognition attainable in the United States for work in applied chemistry.

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WHEN you replace worn-out victims of dezincification with bright, shiny Chase *Antimonial* Admiralty Heat Exchanger Tubes, you're planning on longer service life. Yet you pay no more for *Antimonial* Admiralty than for plain Admiralty tubes!

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*U. S. Pat. No. 2,061,921

Special stocks of ANTIMONIAL ADMIRALTY Condenser Tubes in Chase Warehouses at BALTIMORE, NEW ORLEANS LOS ANGELES, SAN FRANCISCO also carried by— Vinson Supply Co., TULSA Standard Brass & Mfg. Co. HOUSTON

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THE MONTH . . .

troleum Council, declared in a talk at one of the sessions.

"The valves of the industry are wide open now . . . To the very limit of its ability, the oil industry will cooperate with any foreign aid program that may be enacted by Congress, but we must realistically face the fact that we can do just so much," he said. "Consumers of petroleum products must be fully informed that any large scale exports of oil will call for real sacrifice on their part."

Record Attendance

Attendance at this year's American Petroleum Institute's meeting was the highest ever, exceeding last year's previous high by nearly a thousand registrations. Approximately 4700 attended a full agenda of committee meetings and paper presentations. Among those topics discussed by the Division of Refining were a symposium on the evaluation and testing of cracking catalyst and a series of papers on sour crudes.

The tidelands study committee described the Supreme Court's controversial decision as an "unprecedented and unwarranted invasion" of states' right. It further added, "Should Congress fail to nullify the effect of the decision and should it be allowed to stand, the sovereignty of every state in the union will remain in jeopardy."

For developing with his associates the first commercially successful cracking process to make more gasoline out of crude and thereby becoming the father of modern refining technology, Dr. William M. Burton of Cambridge, Mass., retired former president of Standard Oil Company (Indiana) of Chicago, received the American Petroleum Institute's Gold Medal for Distinguished Achievement. This was the Institute's second such medal, the first having been awarded to Henry Ford, founder of the Ford Motor Company, for his work in developing mass-production techniques.

API Officers

With one exception, all officers of the American Petroleum Institute were re-elected. W. Alton Jones, Cities Service Oil Company, New York, replaced J. Brewster Jennings, Socony-Vacuum Oil Company, Inc., as vice president for transportation.

Other officers of the Institute are William R. Boyd, Jr., president; B. J. Majewski, Deep Rock Oil Company, Chicago, vice president for marketing; Robert E. Wilson, Standard Oil Company (Indiana), Chicago, vice president for refining; Jake L. Hamon, Cox & Hamon, Inc., Dallas, vice president for production; L. S. Wescoat, The Pure Oil Company, Chicago, treasurer; and Lacey Walker, Secretary.

Forty-five members were elected to the general committee of the Institute's Division of Refining, and five additional members were appointed by President William R. Boyd, Jr. Those elected members of the refining committee include:

Hines H. Baker, Humble Oil & Refining Company, Houston; George H. Burrus, Anderson-Prichard Oil Company, Oklahoma City; James T-B Bowles, Crown Central Petroleum Corporation, Baltimore; G. A. Davidson, Standard Oil Company of California, San Francisco; C. E. Davis, Shell Oil Company, New York; W. T. Davis, Aetna Oil Company, Louisville; David E. Day, Richfield Oil Corporation, Los Angeles; John C. Day, Western Petroleum Refiners Association, Tulsa; J. H. Dunn, Shamrock Oil and Gas Company, Amarillo; Dewey E. Foster, Wisconsin Oil Refining Company, Sheboygan, Wis.; A. P. Frame, Petroleum Advisers, Inc., New York; Michael Halpern, The Texas Company, New York.

D. P. Hamilton, Root Petroleum Company, Wichita; Basil Hopper, Union Oil Company of California, Los Angeles; H. Y. Hyde, Tide Water Associated Oil Company, San Francisco; F. B. Koentz, Mid-Continent Petroleum Company, Tulsa; J. Porter Langfitt, Pure Oil Company, Chicago; J. A. LaFortune, Warren Petroleum Corporation, Tulsa; R. W. Leslie, Wasatch Refining Company, Salt Lake City; Harry A. Logan, Elk Refining Company, Charleston; R. E. Luton, The Ohio Oil Company, Robinson, Ill.; P. S. Magruder, General Petroleum Corporation, Los Angeles; E. B. McConnell, Standard Oil Company (Ohio), Cleveland; F. W. McCurry, The Derby Oil Company, Wichita; W. E. Moody, Deep rock Oil Corporation, Cushing, Okla.; L. W. Moore, Pan American Refining Corporation, New York; S. B. Mosher, Signal Oil & Gas Company, Los Angeles. John W. Newton, Magnolia Petroleum Company, Beaumont, Texas.

G. G. Oberfell, Phillips Petroleum Company, Bartlesville; Harold G. Osborn, Continental Oil Company, Ponca City; I. A. O'Chaughnessy, Globe Oil & Refining Company, Wichita; M. G. Paulus, Standard Oil Company (Indiana), Chicago; J. B. Rogerson, Lion Oil Refining Company, El Dorado, Ark.; M. R. Ryan, Talco Asphalt & Refining Division of American Liberty Oil Company, Mount Pleasant, Texas; E. L. Shea, Ethyl Corporation, New York; W. F. Sims, Panhandle Refining Company, Wichita Falls.

W. A. Slater, Gulf Oil Corporation, Pittsburgh; C. F. Smith, Standard Oil Company (New Jersey), New York; G. H. Taber, Jr., Sinclair Oil Company, New York; C. S. Teitsworth, Socony-Vacuum Oil Company, Inc., New York; C. H. Thayer, Sun Oil Company, Philadelphia; James W. Vaiden, Skelly Oil Company, Tulsa; John S. Wertz, The Vickers Petroleum Company, Inc., Wichita; W. S. Zehrung, The Pennzoil Company, Oil City, Pennsylvania.

The five members appointed by President Boyd are R. B. Anderson, The Waggoner Estate, Vernon, Texas; Paul G. Blazer, Ashland Oil & Refining Company, Ashland, Ky.; Clyde M. Boggs, Kanotex Refining Company, Wichita; E. J. Henry, The Atlantic Refining Company, Philadelphia; Sidney A. Swensrud, Gulf Oil Corporation, Pittsburgh.

John Bates Heads Sun's New Research Department

Sun Oil Company has created a new independent research department as the initial step toward consolidation of most of the company's scientific activities into a proposed Science Center at Newtown Square, Pa.

Dr. John R. Bates, formerly assistant to Sun's vice president in charge of manufacturing and with the company since 1942, has been appointed director of research and will head the newly-created research department.

DR. JOHN R. BATES

He will be assisted by Dr. J. Bennett Hill, now manager of the development division of the manufacturing department; Dr. John M. Pearson, now director of physical research and development of the production department, and Stewart S. Kurtz, Jr., manager of the experimental division, manufacturing department.

A physical laboratory is being built at Newtown Squire and long-range plans for the science center are under consideration. It is planned eventually to move to the Newtown Squire Center most of the research and experimental work now being done at Sun's Marcus Hook refinery and at its laboratories at Norwood, Pa.

Reorganization of these activities will not affect the geophysical laboratories under the production department in Dallas and Beaumont, Texas. Certain technical service groups will continue their scientific work at Marcus Hook and Toledo refineries together with Sun's automotive laboratory, it was explained.

Sun's proposed science center will be developed along institutional lines. It is situated in a non-industrial area free of smoke, fumes and vibration, about 15 miles from Sun's Marcus Hook refinery.

API Approves Expanded Public Relations Plan

Plans to raise $1,855,000 for an expanded 1948 public relations program for the oil industry were approved by the board of directors of the American Petroleum Institute at the November meeting in Chicago.

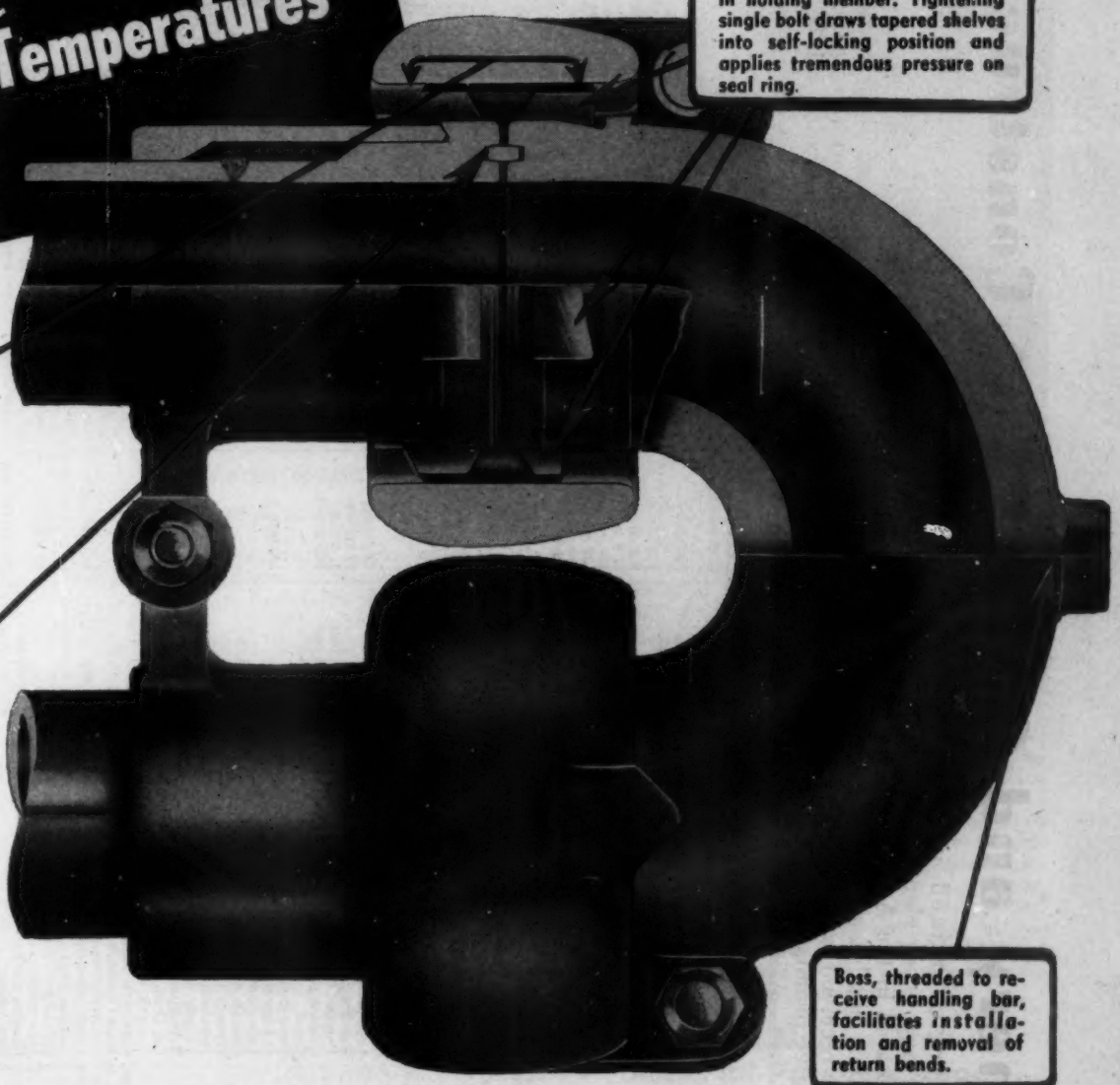
The program includes continuation throughout 1948, upon an expanded basis, of national advertising which was inaugurated last August.

The new program also contemplates an expansion of the number of district field offices and an increase in the number of field representatives.

Commenced a year ago, the oil public relations program during this year was devoted in large part to enlisting the 1.5

Why Unibolt Furnace Fittings Seal Even The Lightest Gases At Extreme Temperatures

Inside Story



9-degree tapered shelves made integral with return bend body, interlock with identical shelves in holding member. Tightening single bolt draws tapered shelves into self-locking position and applies tremendous pressure on seal ring.

ort distance between ints of support minimizes ntraction and expansion. itial seat pressure main- ined uniformly throughout mperature cycle.

Loose metal seal ring seats in 18-degree included angle in ring grooves, providing additional multiplication of pressure applied to the bolt (Total: 20 to 1). Ring and grooves machined to such close tolerances that distorting pressure need not be applied to effect seal.

Boss, threaded to receive handling bar, facilitates installation and removal of return bends.

The fact that UNIBOLT Streamlined Return Bends may be depended upon to provide a dependable seal of light gases under elevated temperatures has been proven on a number of installations where furnace outlet temperatures range up to 1600 degrees F. *This positive seal is accomplished by tightening a single bolt . . .* No rolled-in tubes, no seal welding, no tapered plug closures, no lapping to make a seal, no need for steam in the header boxes. They permit faster "heading up," replacement of a single tube and returning any return bend body to any two tubes. Completely interchangeable parts assure low-cost maintenance.

For complete details, write for Bulletin No. R-42.

NEW TYPE TUBE ALIGNING - ANCHOR CLAMP

The UNIBOLT Tube Aligning-Anchor Clamp serves two purposes: It provides a rigid support for the furnace tubes when the return bend body is removed for tube inspection. Heretofore, the tubes might move out of position when the bend was removed. The clamp also serves as an anchor for the holding member, thus eliminating the split ring which was formerly provided for this purpose. The clamp remains in position until it is desired to replace the tubes.

THORNHILL-CRAVER COMPANY

HOUSTON UNIBOLT TEXAS

Refinery and Other Plant Construction

FURTHER revisions are contained in this month's "box score" detailing refinery and other plant construction projects. The "box score" first appeared in the August issue of PETROLEUM REFINER and since that time has become very popular. A number of suggestions have been received and these, together with other data sent in by readers, give promise toward making the compilation more and more complete as the industry becomes advised as to its purpose and intent.

Companies whose projects are not listed are urged to: 1) send in data on units not listed; 2) furnish details which are missing in the current tabulation and 3) report from time to time on progress of such work. Such helpful cooperation will be very much appreciated.

COMPANY	Plant Site	Project	Daily Capacity	Estimated Cost	Status	Probable Completion	Licenser	Engineering	Contractor
Agwi Pet. Corp.	Fawley, Eng.		120,000 bbls.	$140 million	Projected	1952			
Atlantic Ref. Co.	Philadelphia, Pa.	Propane, lubes			Under Constr.	Early '48	JUIK, Texaco	Badger	Badger
Barceo Oil Co.	Wichita, Kans.	Cat desul. Reforming	7,500 bbls.	$500,000	Under Constr.	Early '48	Phillips	Koch Eng. Co.	Koch Eng.
Barnsdall Oil Co.	Benton, Louisiana	Gasoline plant			Under Constr.				Lummus
Caltex Pet. Maatschappij N.V.	Rotterdam, Holland	Crude plant	20,000	$18 million	Projected	Early '50		McKee	McKee
Carter Oil Co.	Billings, Mont.	Crude cat crk.	30,000 bbls.	$20 million	Under Constr.	3rd Qtr. 1949		U.O.P., S.O.D.	Fluor Corp.
Carthage Corporation	Carthage Field	Gasoline Plant	100,000 mcf	$3.5 million	Under Constr.	Jan. 1, '49		Hudson	Hundson
Carthage Hydrocol, Inc.	Brownsville, Texas	Hydrocol	64,000 mcf	$19 million	Under Constr.	1948	Hydro-Res. Inc.,Texaco	McKee	McKee
Citcon Corp*	Lake Charles, La.	Lube Plant	6,000 bbls.	$30 million	Under Constr.		Texaco		Lummus, Max. B. Miller
Cities Service Oil Co. (Delaware)	East Chicago, Illinois	Modernize plant		$18 million	Under Constr.	Late '48	M. W. Kellogg	Kellogg	Kellogg
Cities Service Oil Co. (Delaware)	East Chicago, Illinois	Laboratory		$300,000	Under Constr.	Spring '48		Cities Service	Sumner-Sollitt
Consolidated Ref. Ltd.	Haifa, Palestine	Crude Plant	124,000 bbls.			By 1950	Lummus, Kellogg	Lummus	Badger
Continental Oil Co.	Billings, Mont.	Cat cracking	7,500 bbls.	$8.5 million	Contract signed	1949	U.O.P.	U.O.P.	J & L Supply
Continental Oil Co.	Denver, Colorado	Crude, cat cracking	7,500 bbls.	$4 million	Under Constr.	Fall 1948	Houdry Corp. & U.O.P.	Con. Oil Co., Lummus	Lummus
Creole Pet. Corp.	Amuay Bay, Venezuela	Cat cracking	60,000 bbls.	$70 million	Under Constr.	1950			
Egyptian Government	Egypt	Crude plant						Lummus	Lummus
Ethyl Corp.	Baton Rouge, La.	Manu. facilities		$20 million	Announcement				
Gray-Wolfe Co.	Montgomery Co.	Gasoline plant	13,000 mcf	$2 million	Under Constr.	Jan. 1949		O. L. Olsen	O. L. Olsen
Gulf Oil Corp.	Crane Co., Texas	Gasoline Plant	23,000 mcf						
Gulf Oil Corp.	Port Arthur, Texas	Crude, Poly-Perco			Under Constr.	Early '48	Gulf-Phillips	Badger, Lummus	Badger, Lummus
Gulf Oil Corp.	Philadelphia, Pa.	Polyform			Under Constr.		Gulf-Phillips	Lummus	Lummus
Gulf Ref. Co.	Toledo, Ohio	Crude unit			Under Constr.	Earl '49		Badger	Badger
Humble Oil & Refg. Co.	Baytown, Texas	Lube plant		$5 million	Under Constr.	Early '48	Texaco	Badger	Badger
Humble Oil & Refg. Co.	East Texas	Gasoline plant		$2 million	Under Constr.			Fish Eng. Corp.	
Humble Oil & Refg. Co.	Katy, Texas	Propane, Butane Extraction	19,000 bbls.	$7 million	Contract let				Stearns-Roger
Humble Oil & Refg. Co.	Refugio, Texas	Gasoline plant	35,000 mcf	$2.5 million	Under Constr.	1948			Gaso. Plant
Industrie Chimiche Italiane Petroli	Mantau, Italy	Crude	4,500 bbls.						
Jefferson Chemical Co.	Port Neches, Texas	Petro-chem			Under Constr.	1948		Jefferson	Jeff'n, Badger, Lummus
Kanotex Refg. Co.	Arkansas City, Kans.	Cat cracking	4,200	500,000	Contract let	Early '48	Phillips	Koch Eng.	Koch Eng.
Magnolia Pet. Co.	Beaumont, Texas	Lube plant			Under Constr.	Jan. 1, '49	Texaco	Lummus	Lummus, Kellogg
Magnolia Pet. Co.	Seeligson Field, Texas	Gasoline plant	90,000 mcf			Late '48		Hudson	
Fred M. Manning, Inc.	Stephens Co., Texas	Pressure Maintenance	20,000 mcf	$500,000	Contract let	Jan. 48'		Fish Eng. Corp.	
McCarthy Chemical Co.	Winnie, Texas	Petro-chem, Cat cracking	5,000 bbls.	$3 million	Under Constr.		Phillips	Fish Eng. Corp.	
Mid-Continent Pet. Co.	Tulsa, Okla.	Vac still, cat cracking	45,000 bbls.	$8.5 million	Contract let	June '48	Kellogg	Kellogg	Kellogg
Monsanto Chemical Co.	Texas City, Texas	Rebuilding Petro-chem.		$6 million	Contract let				Bellows Co., Leon Con. Co.
National Co-Op. Ref. Assn.	McPherson, Kansas	Cat Desul and Ref. unit	5,000	$400,000	Under Constr.	Late '47	Phillips	Koch Eng.	Koch Eng
Ohio Oil Co.	Robinson, Illinois	Vac unit, cat cracking	27,500 bbls.	$15 million	Under Constr.	Late '48	U.O.P.	McKee	McKee, Pritchard
Petro Carbon, Ltd.	Partington, England	Cat Cracker, Lube and Reform	20,000 bbls.						
Petro Chemical, Ltd.	Partington, England	Petro Chemicals							
Petrol Terminals Corp.	Texas City, Texas	Cat cracker (recondition)			Under Constr.	Late '47			Cat. Constr.
Petroleos Mexicanos	Salamanca, Mexico	Crude, gasoil cracking	30,000	$12 million	Contract let	1949	U.O.P.	McKee	McKee
Plomo Co.	Nueces Co., Texas	Gasoline plant			Awaiting Appr'l				
Porto Marghera Rfg.	Venice, Italy	Enlarging							
Pure Oil Co.	Smith's Bluff, Texas	Lube plant		$9 million	Contract let		Texaco	Lummus	Lummus
Richardson and Bass	Keystone Field, Texas	Gasoline plant	70,000 mcf		Constr. approved	10-1-48		J & L Supply	J & L Supply
Shell Chemical Co.	Thornton, England	Petro-chem.		Several millions	Under Constr.				
Shell Chemical Co.	Shell Haven, England	Petro-chem.							
Shell-Chemical Co.	Stanlow, England	Petro-chem.							
Shell Chemical Co.	Houston, Texas	Glycol		Part $25 mln.	Under Constr.	Early '48			Kellogg
Shell Chemical Co.	Shell Point, California	Syn. Ammonia				Early '48			
Shell Oil Co.	Notrees, Texas	Gasoline plant	45,000 mcf	$4 million	Under Constr.	March '48		Brown & Root	Brown & Root
Shell Oil Co.	Houston, Texas	Lube unit		$10 million	Under Constr.		Texaco	Lummus	Lummus
Shell Oil Co.	Denver City, Texas	Gasoline plant	24,000 mcf		Under Constr.	1948			Fluor Corp.
Shell Oil Co.(Venezuela)	Cardon, Venezuela	Complete refinery			Under Constr.			Lummus	Lummus
Sinclair Ref. Co.	East Chicago, Illinois	Lube plant						Lummus	Lummus
Socony-Vacuum Oil Co., Inc.	Paulsboro, New Jersey	Crude, Lube plant			Under Constr.	Jan. 1, 1949	Texaco	Badger	Badger
Socony-Vacuum Oil Co., Inc.	Casper, Wyoming	Cat cracking	7,500		Under Constr.	Jan. 1, 1949			
Socony-Vacuum Oil Co., Inc.	East St. Louis, Illinois	Thermal Cracker and distillation			Projected	Jan. 1, 1949			
Socony-Vacuum Oil Co., Inc.	Augusta, Kansas	Cat cracking	10,000			Jan. 1, 1949		Cat. Constr. Co.	Cat. Constr. Co.
Stanolind Oil & Gas Co.	Elk Basin, Wyoming	Gasoline plant	12,000 mcf		Approved			Stone & Webster	Stone & Webster
Stanolind Oil & Gas Co.	Slaughter, Texas	Gasoline plant	90,000 mcf		Approved			C. F. Braun	C. F. Braun
Stanolind Oil & Gas Co.	Hugoton Field, Kansas	Gasoline plant	150 mcf		Approved			J & L Supply	J & L Supply
Stanolind Oil & Gas Co.	Hugoton Field, Kansas	Gas syn plant	100,000 mcf		Announced			Hydro Res.	Kellogg, Foster Wheeler, Stone & Webster
Stanolind Oil & Gas Co.	Hockley County, Texas	Gasoline Plant							
Stanolind Oil & Gas Co.	Hugoton Field, Kansas	Chemical plant	130,000 lbs./yr.		Announced			Badger	Badger
Stanolind Oil & Gas Co.	Brownsville, Texas	Chemical plant	130,000 lbs./yr.		Announced			Badger	Badger
Stanolind Oil & Gas Co.	Pettus, Texas	Cycling plant	159,000 mcf		Projected	1949		J & L Supply	J & L Supply
Standard Oil Co. (Indiana)	Sugar Creek, Missouri	Cat cracking	25,000 bbls.						

Stanolind Oil & Gas Co.	Hockley County, Texas	Gasoline Plant	130,000 lbs./yr.		Announced			Badger	Badger
Stanolind Oil & Gas Co.	Hugoton Field, Kansas	Chemical plant	130,000 lbs./yr.		Announced			Badger	Badger
Stanolind Oil & Gas Co.	Brownsville, Texas	Chemical plant	159,000 mcf		Projected	1949		J & L Supply	J & L Supply
Stanolind Oil & Gas Co.	Pettus, Texas	Cycling plant	25,000 bbls.			Early '48			Kellogg
Standard Oil Co. (Indiana)	Sugar Creek, Missouri	Cat cracking	25,000 bbls.			Early '48			Kellogg, Lummus
Standard Oil Co. (Indiana)	Whiting, Indiana	Cat cracking	90,000 bbls.		Contract let	1st Qtr. 1949	Texaco		C. F. Braun
Standard Oil Co. (New Jersey)	Linden, New Jersey	Vac. cat cracking, pipe still coker	111,500 bbls.		Under Constr.	Early '48		Badger, S.O.D., C. F. Braun, McKee, Kellogg	Badger, Kellogg, McKee
Standard Oil Co. (New Jersey)	Baton Rouge, Louisiana	Lub, Vac. still, crude, dewax, poly							C. F. Braun
Standard Oil Co. (California)	Richmond, California	Grease	82,000 lbs.	$5.2 million	Under Constr.	Sept. '48	U.O.P.	W. G. Pough	Barrett & Hilp
Standard Oil Co. (Ohio)	Lima, Ohio	Cat cracking	6,000	$10 million	Contract let	Late '49	U.O.P.	KeMee	McKee
Stora Kopparberg	Sweden	Crude, ther cracking						Lummus	Lummus
Sun Oil Co.	Richland Parish, La.	Gasoline plant	70,000 gal.		Under Constr.				Pet. Eng., Inc.
Sunray Oil Co.	Duncan, Oklahoma	Crude plant	20,000 bbls.		Under Constr.	Late '47		U.O.P.	
The Texas Co.	Eagle Pt., New Jersey	Vac. still, cat cracking	40,000 bbls.		Contract let	Late '49		Kellogg, Texaco Co.	Kellogg, Foster Wheeler
The Texas Co.	Lawrenceville, Illinois	Crude and cat cracking	20,000 bbls.		Announced	1950		Foster Wheeler	Kellogg, Foster Wheeler
The Texas Co.	West Tulsa, Oklahoma	Crude and cat cracking	30,000 bbls.		Contract let	1949		Texas Co., Kellogg, Foster Wheeler	Kellogg, Foster Wheeler
The Texas Co.	Port Arthur, Texas	Lube addition	2,000 bbls.		Under Constr.	Late '48		Texas Co., Foster Wheeler	Foster Wheeler
The Texas Co.	Casper, Wyoming	Cat Poly Naphtha reformer, pressure coker	10,000 bbls.		Completed		Kellogg	Kellogg	Kellogg
Winnaboro Ref. Co.	Winnaboro, Texas	Crude plant	3,000	$175,000	Under Constr.	Jan. 1, '48			Mid-Continent
Wood River Oil & Ref. Co.	Hartford, Illinois	Cat cracking, easoli, cracking	8,500	$1 million	Under Constr.	Late '48	Phillips	Koch Eng.	Koch Eng.

NOTE: Badger—E. B. Badger and Sons. Bellows—W. S. Bellows Co. Braun—C. F. Braun Co. Cat. Constr.—Catalytic Construction Company. Fish Eng.—Fish Engineering Corporation. Fluor—Fluor Corporation. Gaso. Plants—Gasoline Construction Company. Houdry—Houdry Corporation. Hudson—Hudson Engineering Corporation. Hydro Res.—Hydrocarbon Research, Inc. Jefferson—Jefferson Chemical Company. J. & L. Supply—Joors & Laughlin Supply Company. Koch Eng.—Koch Engineering Company. Kellogg—M. W. Kellogg Company. Leon Constr.—Leonard Construction Company. Lummus—The Lummus Company. McKee—Arthur G. McKee Company. Phillips—Phillips Petroleum Company. Pet. Eng.—Petroleum Engineers, Inc. Pritchard—J. F. Pritchard & Company. Stearns-Roger—Stearns Roger Manufacturing Company. Texaco—Texaco Development Company. • Jointly owned by Cities Service Oil Co. and Continental Oil Co.

THE MONTH . . .

million men and women within the industry in the fact-telling program. Emphasis next year will be placed on carrying oil's story to stimulate increased action by trade associations and oil companies at the local level.

The board's action was interpreted by William R. Boyd, Jr., API president, as a complete indorsement of the conduct of the industry's current public relations program.

The API board designated its public relations committee, under the chairmanship of John M. Lovejoy, to take over-all charge of the program on behalf of the API, and to establish policies for the conduct of the 1948 program. It is understood that the public relations operating committee of which Franklyn Waltman is chairman, will be authorized by the Lovejoy committee to direct the administration of the program.

Robert H. Colley, chairman of a finance committee, said that his group would meet shortly to discuss raising the fund. All oil companies will be invited to contribute.

The public relations program, already one of the most comprehensive of any ever undertaken by an industry, will be administered and managed by the Fred Eldean Organization, Inc. The advertising agency for the program is Sullivan, Stauffer, Colwell and Bayles, Inc.

Bureau Reports on Pilot Plant Gasification Work

Experimental work just completed by the Bureau of Mines in a pilot plant at Grand Forks, N. D., shows that more than 900 billion tons of lignite in three Western states is a potential source of fuel gas for beneficiating iron ore, making hydrogen, and other industrial uses, James Boyd, Bureau Director, announced last month.

The Grand Forks gasification pilot plant was put into operation following similar experiments on a smaller scale at a pilot plant at Golden, Colo. The current report covers operations at both pilot plants and draws a comparison between small and large pilot plants.

Investigations confirmed results obtained during the operation of the smaller plant, proving that 50 to 90 percent of the carbon in natural lignite can be converted to water gas. The cost of producing industrial gas from lignite by the new process compares favorably with that of other competitive processes.

A copy of Report of Investigations 4128 "Gasification of Lignite and Subbituminous Coal, Progress Report for 1945-46" can be obtained by writing to the Bureau of Mines, Publications Section, 4800 Forbes Street, Pittsburgh, Pa.

R. L. Smith Will Supervise Sinclair's Research Work

Dr. Robert L. Smith has recently been appointed to supervision of the project evaluation division of Sinclair's Research and Development Department at East Chicago. The assignment was made in conjunction with the scheduled expansion of research activities at the new laboratories in Harvey, Ill.

Dr. Smith received his B.S. and Ph.D. degrees in chemical engineering at the University of Michigan. He is the author of eight publications dealing with PVT and thermodynamic relations of gases; methods of design for fractional distillation and absorption; and generalized properties of hydrocarbon mixtures.

For the past 15 months he has been engaged in process development with Sinclair research. His new position involves planning of research programs, project evaluation, data correlation, and development of design methods. Assisting him will be Harry Mayland, also with a background of experience in the field of process development.

Foster Wheeler Names Public Relations Head

John M. West has been appointed director of the public relations division of Foster Wheeler Corporation, New York,

JOHN M. WEST

in charge of advertising, publicity, literature and house organs.

West has been with Foster Wheeler since January, 1940. His previous duties included sales engineering as well as design of equipment.

Stanolind Names Zink Chemical Products Head

Dr. Donald G. Zink, formerly director of technical development of U. S. Industrial Chemicals, Inc., has joined the staff of Stanolind Oil and Gas Company, Tulsa, as superintendent, Chemical Products Department.

Dr. Zink attended Louisiana State University and Johns Hopkins University, receiving his Ph.D. degree in 1934, when he joined U. S. Industrial Chemicals.

The Chemical Products Department has been created by Stanolind to handle all development and marketing in the field of chemicals. Immediate problems concern the large amounts of chemicals which will be produced as by-products in the synthesis of gasoline from natural gas, officials said.

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Program Announced For Regional NGAA Meeting

The program for the Panhandle-Plains regional meeting of the Natural Gasoline Association of America, scheduled December 12 at the Herring Hotel, Amarillo, was announced late last month and highlights a morning session with papers of special interest to gasoline plant operators and an afternoon session devoted to broad general interests of the natural gasoline and related industries of the area.

R. C. Kay, independent operator of Pampa, Texas, is chairman of the arrangements committee and B. R. Carney of Shamrock Oil and Gas Corporation, chairman of the program committee. Papers include:

"Introduction," N.G.A.A. President, C. R. Williams, The Chicago Corporation, Corpus Christi; "Water Treatment for Gasoline Plants," J. C. Hutchinson, Shell Oil Company, Great Bend; Prepared Discussion, J. S. Connors, Phillips Petroleum Company, Borger; "Maintenance and Operation of Plug Valves," Roy Bush, Nordstrom Valve Company, Tulsa; "Plant Operations for Maximum Recovery of LPG," J. A. Craig, Skelly Oil Company, Skellytown, Texas; Prepared Discussion, B. O. Bentley, Hagy, Harrington and Marsh, White Deer, Texas; "Changing Times in the Natural Gas and Gasoline Industry," L. R. Hagy, mayor of Amarillo; "Carbon Black in the Texas Panhandle," Ira Williams, J. M. Huber Corporation, Borger; "Progress of the Panhandle Gas Field," E. O. Thompson, Texas Railroad Commission, Austin: and "The Good Oil Days of Natural Gasoline," S. C. Carney, Phillips Petroleum Company, Bartlesville, and J. J. Hastings.

Another such regional meeting will be held in Corpus Christi, Texas, in late January or early February.

Koppers Company Acquires Manufacturers Chemical Corp.

Koppers Company, Inc., Pittsburgh, last month acquired all stock of the Manufacturers Chemical Corporation and its sales subsidiary, the Chemaco Corporation, both of Berkeley Heights, N. J. Koppers Chemical Division which will operate the newly acquired plant will enable Koppers to serve to large and small users of various plastic molding compounds, inasmuch as Manufacturers is engaged in processing and Chemaco is engaged in the sale of compounds made of polystyrene, cellulose acetate, and ethyl cellulose.

Lee Joins Pennsalt

William M. Lee has joined the research and development department of Pennsylvania Salt Manufacturing Company as supervisor of the special products division. His experience includes wartime service as chief of the chemicals and plastics section of the office of the Quartermaster General.

Lee is a member of the American Chemical Society, Theta Chi, and Alpha Chi Sigma.

L. F. McCOLLUM

DAN MORAN

McCollum Succeeds Moran As Continental President

Leonard F. McCollum, coordinator of producing activities for Standard Oil Company (N. J.), became president of Continental Oil Company December 1, succeeding Dan J. Moran, who will continue with the firm in a consulting capacity.

McCollum is succeeded by John W. Brice, for the last 20 years engaged in exploration and producing activities for Jersey Standard and affiliates.

McCollum, born in 1902 in Tennessee, is a 1925 geologist graduate of the University of Texas, did early work for Humble Oil & Refining Company and The Carter Oil Company, of which latter company he came to be executive vice president. He has been with Jersey since 1943.

Moran, born in 1888 in Ohio, joined Buckeye Pipe Line Company in 1905, took a degree in 1907 from the University of Dayton, did graduate work at the Case School of Applied Science. He joined The Texas Company and in 1921, becoming its vice president and general manager in Mexico where he built two refineries and pipe line facilities. He later returned to the U. S. and served as Texaco vice president and executive committee member. He became president of Marland Oil Company in 1928, reorganized that firm and changed the name to Continental Oil Company.

Brice, a native (1903) Texan, is another University of Texas geology graduate (1927), worked with several Jersey affiliates and in 1945 went to New York as assistant coordinator of producing activities for Jersey.

Boatwright Is Promoted By Standard of Indiana

John W. Boatwright has been appointed general manager of the distribution economics department of Standard Oil Company (Indiana). He formerly headed the combined sales research and budgeting departments. In his new position he will be responsible for all functions relating to sales budgets, general economic studies, problems of product supply, consumer research studies, economic evaluation studies and profit and loss analyses previously performed in the sales research and coordinating departments.

D. S. Warning has been appointed manager of the commercial research division of the distribution economics department, and R. M. Eberly manager of the coordinating division of the distribution economics department.

Dr. Chaney Will Direct Gas Production Research

Dr. Newcomb K. Chaney, for the past 12 years director of research for The United Gas Improvement Company, Philadelphia, on November 1 became director of the gas production research program of the American Gas Association. Dr. Chaney takes charge of the work previously directed by Edwin L. Hall who has been appointed director of the Association's Testing Laboratories at Cleveland and Los Angeles.

The association is engaged in a continuing program of research to find better and more economical ways to produce gas, projects being under way at the AGA Laboratories, in utility plants, and at a number of universities, scientific and governmental institutions outside the industry. Dr. Chaney will coordinate and guide these activities.

Socony Vacuum Group Sets Up New Sigma Xi Chapter

More than 40 members of the staff of the research and development department of Socony-Vacuum Oil Company, Inc., at Paulsboro, N. J., participated November 17 in installation of a new Socony-Vacuum unit of the Society of Sigma Xi, national organization for the encouragement of scientific research.

Officers of the new unit include Dr. D. E. Badertscher, Woodbury, N. J., president; Dr. Ronald Bulkley, Haddon Heights, N. J., vice president, and Dr. Howard Hartough, Pitman, N. J., secretary-treasurer.

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Northwestern Dedicates Ipatieff Laboratory

The Ipatieff Catalytic and High-Pressure Laboratory for Chemical Research has been formally opened by Northwestern University, Chicago. Major functions of the laboratory will be to introduce to chemistry and chemical engineering students the methods of scientific research and catalysis operation encountered in industry today. The teaching program will include lectures and experiments whereby improvements in technique and equipment will be introduced as they are developed. The new teaching laboratory will be available to both graduate and undergraduate students.

Dr. Vladimir N. Ipatieff, in whose honor the laboratory was named, will be general director with Dr. Herman Pines of Northwestern University as associate director.

A Russian-born scientist, Dr. Ipatieff joined the faculty of the university in 1931 and in recent years has devoted a great deal of his time to the training of young scientists and aiding the cause of democracy through chemical research.

Majewski Withdraws Plan For Scheduling Oil Supply

B. L. Majewski's proposal for scheduling of crude oil, transportation and product supplies (PETROLEUM REFINER, November) was withdrawn early in November by its author who thus acquiesced to widespread industry opposition to the plan. Majewski in a letter October 15 to Max W. Ball, director of OGD, had suggested methods which many construed as the first step toward government regulation of oil.

Majewski, president of Deep Rock Petroleum Company, said "in bowing to superior majority judgment, I acknowledge the functioning of the democratic processes without yielding one iota in my determination to see to it that the industry will fulfill its responsibility to the consumer and to the government."

Krug Calls Upon Nation To Conserve Oil Supply

Urging nationwide vigilance to prevent waste of fuel and energy, Secretary of the Interior J. A. Krug said last week that shortages in preferred fuels unquestionably will be felt in some areas this winter owing to immense demands and inadequate transportation facilities.

Transportation remains the No. 1 bottleneck, he said, citing the prevailing shortages of railroad coal cars and tank cars and tank ships, and a lack of sufficient pipe lines to supply all requests for natural gas. Excerpts:

For light fuel oils including distillate used in home heating, it is estimated that domestic demand in 1948 will be some 50 percent greater than in 1945. Intensified by continued scarcities in coal and other competing fuels, demand increased about 9 percent in 1946, will gain at least 20 percent in 1947, and is expected to continue to increase rapidly in 1948. Reasons: 1) a much higher domestic demand than anticipated; 2) world shortage of refinery capacity and products which forestalls increasing imports and keeps exports at a high level; and 3) lack of adequate transportation facilities—new tank cars, pipe lines and tankers—to distribute products properly in the United States.

With the winter heating season beginning, Secretary Krug pledged the full technical services of the Department of the Interior in aiding consumers to attain utmost efficiency from available supplies.

Two New Cat Crackers Planned by Gulf in Ohio

Plans for improvements to Gulf Oil Corporation's refineries at Cincinnati and Toledo, Ohio, announced last month, calls for large new catalytic cracking units (like the one pictured above) at both plants. In addition, production at the Toledo plant will be increased 25 percent. Contracts have been let and major construction is scheduled to start in early spring, and completed in 1949. The program is estimated to cost in excess of $20 million. The new catalytic cracking unit to be erected at Cincinnati will largely replace present thermal cracking installations, which will be converted mainly to primary crude distillation. A new polymerization unit also will be built. The machine shop and other mechanical shops will be modernized and combined in one building. Additions will be made to the laboratory and office building. At Toledo, installation of a new catalytic refining unit will increase production capacity of gasolines and domestic fuels by 25 percent. A new primary crude distillation unit will be included. In addition, a modern laboratory will be built and all mechanical shops will be centralized under one roof. Chicago Bridge and Iron Company has contracted for relocating and rebuilding tanks at Cincinnati, where the site is now being cleared. Contractors have begun road, track and levee work at Toledo. E. B. Badger & Sons Company and The M. W. Kellogg Company will perform major construction at both refineries.

Jersey Standard Proceeds With Italian Company Deal

Restoration of the war-damaged oil refinery at Livorno, Italy (prewar capacity, 6000 barrels daily), is one of the aims of negotiations being conducted by Standard Oil Company (New Jersey) with the Italian government. Standard also seeks an exploration concession in the Po Valley.

Another refinery at Livorno (7000 barrels daily capacity) was not damaged and is currently operating.

Carlson Dies

Leonard O. Carlson, 54, president of Erie Meter Systems, Inc., and one of its co-founders, died November 9 at his home in Erie, Pa.

Paige Named New Head Of Ethyl's Fleet Section

Edwin C. Paige has been named head of the fleet section of the technical service division of Ethyl Corporation. This division functions as part of the Ethyl research laboratories in Detroit, serves as a liaison group between the laboratories and the petroleum, automotive, and agricultural machinery industries.

Paige succeeds Errol J. Gay, who resigned to open a Detroit office as private consultant to petroleum and automotive companies on problems relating to engines, fuels, and lubricants.

Paige joined Ethyl Corporation in 1930, two years after his graduation from the University of Maryland with a civil engineering degree.

Huge Towboat Is Launched For Standard of Louisiana

The "Big Mamma" of all towboats, the "Esso Louisiana," which will replace the largest towboat ever built, the "Sprague," in the service of Standard Oil Company of New Jersey, was launched last month from the Dravo Corporation ways, Pittsburgh, Pa.

An interesting paragon of the improvements in propulsion efficiency is the fact that although the "Sprague" is 110 feet longer and 26 feet wider than

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the new diesel propelled twin screw "Esso Louisiana" the smaller vessel is expected to out-perform the "Sprague" in every respect.

Mrs. M. W. Boyer, of Baton Rouge, La., was the sponsor. Mrs. Boyer's husband is vice president and director of Standard Oil Company of New Jersey, which will operate the towboat between Baton Rouge and Louisville.

The 166-foot "Esso Louisiana" is one of two sisterships being built for Standard of New Jersey. The boat, manned by two 1000 hp diesels, will be one of the first vessels of her type on the inland waterways to carry radar and ship-to-shore telephone. She is designed to handle 12 oil barges of 9000 barrels capacity, a total of 108,000 barrels at an average speed of about 4 mph.

Cosden Promotes Orme

Douglas Orme has been elected vice president in charge of traffic of Cosden Petroleum Corporation effective November 17.

Orme joined Cosden in 1929 and has served in the Traffic Department since that time. He was promoted to traffic manager in December 1935. He has resided in Big Spring, Texas, since December, 1939, when the offices of the traffic department were moved from Fort Worth.

Course on Waxes

Leo Ivanovszky has advised THE REFINER that he will give a course of eight postgraduate lectures on "The Modern Chemistry and Technology of Waxes" at Chelsea Polytechnic, London, England, on Friday evenings commencing January 16, 1948. The first course, offered last year by Ivanovszky, was attended by some 60 persons, mostly postgraduate students, research workers and industrial chemists.

Nation's Transport System Handles Record Oil Volume

The nation's oil transport system is now geared to handle an average of 11.5 million barrels of petroleum and petroleum products a day, an all-time record volume, according to a survey presented in *The Lamp*, publication of Standard Oil Company (New Jersey). But it is pointed out, even this amount is insufficient to meet today's unprecedented consumer demands.

During the first six months of this year, pipe lines, ocean tankers, railroad tank cars, barges and trucks moved an average of 5,720,000 barrels of petroleum a day from sources of supply to refineries and terminals. In addition, 5,780,000 barrels of refined products a day went from refineries and terminals to consumers. This amounted to the movement of 11,500,000 barrels of oil and oil products a day.

"Before the war, this transportation system was able to keep pace with the demand for products," *The Lamp* noted. "Not only did the demand increase during the war, however, but it is continuously reaching new record levels. Use of petroleum products has risen from a rate of 469 gallons for every person in the United States in 1941 to a present annual rate of 574 gallons.

"To help meet these demands, the oil industry is spending $440 million this year and next on transportation expansion alone.

"Pipe lines are first on the list. During the first six months of this year, for example, the nation's oil wells produced an average of 4,916,000 barrels a day. Of this amount, the industry's 133,000 miles of pipe lines carried about 4,734,000 barrels a day. But new lines are needed, and up to last July, the industry had made plans for 7600 miles of new crude oil pipe lines.

"Improvements also are being made in barge equipment. Last year there were 1353 tank barges on the Mississippi River system with a total capacity of nearly 10 million barrels. Equipment capable of carrying an additional million barrels has been added since then. Despite this, there still are not enough barges to meet the steadily increasing demand.

"Of the 118,000 railroad tank cars now available to the industry, many of them are overdue for retirement, and shortages exist in all types. Orders have been placed with car builders, but without much likelihood of getting the orders filled in the near future.

"In the case of tank trucks, there are about 200,000 of them, their number having been increased some 22 percent since the war ended.

"Current expansion of the transport system is proceeding as rapidly as construction materials become available."

Named Assistant Manager Of Chemical Division

Dan M. Rugg, vice president and general manager of the chemical division of Koppers Company, Inc., has announced appointment of B. J. C. van der Hoeven as assistant general manager of the division, effective at once. The position is a newly created one in Koppers chemical division.

During the war van der Hoeven was manager of operations of the Kobuta, Pa., styrene and butadiene plant which was built and operated by Koppers for the government. Since early last year, he has been chief of the company's chemical engineering department, engineering and construction division. In the latter position he has been responsible for the design, selection of equipment, or erection of several new chemical plants built or acquired by Koppers in the past year.

Van der Hoeven, a native of The Netherlands, joined the research department of Koppers in 1926.

Commercial and College Laboratories Are Listed

A complete listing of commercial and university testing and research laboratories throughout the country, together with indications of the type of commodities tested, has been compiled by the National Bureau of Standards. This pamphlet is available from the Government Printing Office, Washington 25, D. C., as NBS Miscellaneous Publication M187 entitled *Directory of Commercial and College Laboratories* (30 cents per copy).

The Bureau does not make tests for private individuals if other laboratories can do the work with the required accuracy and since inquiries regarding such services are often received, this classified list was assembled to inform interested persons of the location of other testing facilities.

ACS Readers Select Outstanding Engineers

The "ten ablest chemists and chemical engineers now working in the United States" in the petroleum field were listed last month in The Chemical Bulletin of the Chicago Section of the American Chemical Society as follows:

Gustav Egloff, Universal Oil Products Company; M. R. Fenske, Penn State University; F. E. Frey, Phillips Petroleum Company; V. N. Ipatieff, Northwestern University; P. C. Keith, Hydrocarbon Research, Inc.; E. V. Murphree, Standard Oil Company (New Jersey); T. H. Rogers, Standard Oil Company (Indiana); F. D. Rossini, Bureau of Standards; Louis Schmerling, Universal

Trends of Operations and Changes in Stocks

Figures on crude stocks are from Bureau of Mines weekly reports; all others from American Petroleum Institute weekly reports, which are estimates on Bureau of Mines basis.

(All figures in thousands of barrels—add 000)

Trends in Week Ended	Crude Oil			Gasoline		Gasoil and Distillate		Residual Fuel	
	Production Daily	Runs to Stills Daily	Stocks Week End	Production Weekly	Stocks Week End	Production Weekly	Stocks Week End	Production Weekly	Stocks Week End
1946:									
January 26	4,626	4,553	220,544	13,622	101,737	5,720	29,498	8,411	39,722
March 2	4,726	4,779	229,430	13,871	104,462	5,888	25,148	8,634	38,441
March 30	4,425	4,684	224,994	13,896	104,715	5,337	28,240	8,738	37,746
April 27	4,672	4,685	224,443	14,228	99,631	5,568	30,466	9,204	39,404
May 25	4,759	4,857	222,214	14,322	95,769	5,463	32,973	8,908	43,368
June 29	4,957	4,854	223,883	14,500	92,333	5,325	37,762	8,828	46,447
July 27	4,926	4,896	223,756	14,535	88,626	5,817	44,316	8,217	49,517
August 31	4,833	4,901	225,672	15,405	86,813	5,374	53,858	8,357	53,037
September 28	4,778	4,829	221,903	14,538	85,854	5,450	57,903	8,158	56,914
October 26	4,730	4,758	221,184	14,863	84,623	5,710	65,499	7,728	60,872
November 30	4,795	4,707	225,119	15,145	88,371	5,228	66,062	7,672	58,647
December 28	4,713	4,968	226,111	15,604	93,126	5,931	58,941	8,181	53,427
1947:									
January 25	4,672	4,820	221,655	14,624	99,801	5,630	50,357	8,224	48,558
February 22	4,786	4,860	224,580	14,668	103,904	5,929	40,739	8,532	44,919
March 29	4,865	4,843	225,720	14,396	107,576	5,969	32,737	8,668	43,364
April 26	4,930	4,725	234,051	14,213	103,860	5,435	32,286	8,186	42,668
May 31	5,024	5,000	239,370	14,709	95,876	5,732	36,032	8,910	45,224
June 28	5,109	5,093	236,221	16,070	91,806	5,606	41,721	8,298	48,490
July 26	5,064	5,162	231,357	16,142	85,812	5,708	47,097	8,738	52,497
August 30	5,157	5,209	228,281	16,753	84,105	5,919	54,808	3,962	55,242
September 27	5,196	5,290	224,994	16,501	82,472	6,136	60,406	8,580	56,777
October 25	5,295	5,165	224,308	16,529	82,008	6,638	62,609	8,666	57,419
November 29	5,257	5,309	*224,005	16,667	86,319	6,382	61,988	8,405	56,747

* Stocks, November 22.

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Oil Products Company, and Robert E. Wilson, Standard Oil Company (Indiana).

Such choices also were made by The Chemical Bulletin in 19 other specialized fields, all the selections being the result of a reader poll, it was explained.

NACE 1948 Conference, Exhibit Set for St. Louis

The fourth annual conference and exhibition of the National Association of Corrosion Engineers will be held April 5-8, 1948, at the Jefferson Hotel, St. Louis. The technical program will include symposia on cathodic protection, chemical industry, communications, electrical industry, gas industry, general industry, oil industry, protective coatings for metals, salt water corrosion and water industry.

The annual banquet will be held Wednesday evening, April 7.

Crown Central Purchases Plant Units at Pasadena

A Pasadena, Texas, 100-octane gasoline refinery, operated for the government during the war by Crown Central Petroleum Corporation, has been sold to that corporation for $2,300,000.

The plant, on a nine-acre site, includes: a thermofor catalytic cracker of nominal 10,000-barrel-per-day capacity, a treater of nominal 6000-barrel-per-day capacity, an HF alkylation unit with a capacity of 3500 barrels per day, and a butane isomerization unit, together with various auxiliaries and utilities.

Plans Discussed For 1948 Instrument Society Meet

The Philadelphia committee of the Instrument Society of America comprising E. J. Grace, Jr., general chairman, C. W. Wood, M. H. White, F. H. Trapnell, national vice president, J. Johnston, Jr., G. N. Ehly, L. H. Allen, W. J. Gallagher, J. C. Whiddett, P. Putnam, E. A. Adler, Stuart Rogers and Richard Rimbach, national executive secretary, has held its first meeting to make arrangements for the Third Instrument Conference and Exhibit to be held in Convention Hall, Philadelphia, September 13-17, 1948.

The exhibit will be known as the "American Instrument Fair." A highlight will be an instruments historical display on the Arena stage.

Sessions will be held by the Instrument Society of America and a number of co-sponsoring societies.

Keffer Dies at Ashland

William H. Keffer, 63, vice president and director of Ashland Oil & Refining Company, died November 17 in Ashland, Ky., following a heart attack. He had been ill for several days. Funeral services were held November 19 in Ashland.

Keffer had spent his entire life in the oil industry and at the time of his death was in charge of pipe lines and crude oil production for the Ashland Oil & Refining Company.

W. HENRY LINTON

Linton Named Assistant To Clarence Thayer by Sun

W. Henry Linton has been named assistant to Clarence H. Thayer, vice president in charge of manufacturing, Sun Oil Company. Linton, formerly manager of the refinery accounting department, will serve as an administrative assistant.

In 1919 Linton joined Sun Shipbuilding and Dry Dock Company and three years later entered the accounting department of Sun Oil at Philadelphia. In 1926 he was placed in charge of the yield department at Sun's Marcus Hook refinery, continuing there until 1931 when he returned to the Philadelphia office.

From 1937 to 1942 Linton served as office manager and chief accountant at Marcus Hook. He became manager of refinery accounting in 1942.

Greenwood to Deer Park

Shell Oil Company, Inc., has announced that Fred H. Greenwood of Alton, Ill., has joined the staff of its Deer Park refinery as technologist in the lubricating oil plant.

Greenwood, a graduate of Illinois Institute of Technology, was formerly employed as technologist in the company's Wood River, Ill., refinery.

Ahlberg Joins Davison

J. Elston Ahlberg has joined the research staff of The Davison Chemical Corporation, Baltimore, Md. Dr. Ahlberg's principal work in the oil industry has been development of methods of manufacturing cracking catalysts. During the past eleven years he was associated with Universal Oil Products Company, Chicago.

Don Carroll Named

Don Carroll, former staff member of *World Oil* (The Oil Weekly) has been named assistant to Max Ball, director of the Oil and Gas Division of the Department of the Interior.

George Reese Named Chief Chemist at Marcus Hook

George E. Reese has been named chief chemist at Sun Oil Company's Marcus Hook refinery succeeding Edward M. Hughes, chief chemist since 1919, who has been appointed to the newly-created position of consultant chemist. Theodore R. McClure will succeed Reese as assistant chief chemist.

Reese joined Sun Oil Company as a refinery laboratory chemist in 1924 shortly after graduating from Gettysburg College with a B.S. degree in chemistry. He has been assistant chief chemist since 1939.

Hughes, a 1909 B.S. in chemistry graduate from the University of Pennsylvania, joined Sun in 1911. In 1916 he developed a process for making naphthenic soap from the caustic treatment of lubricating oils.

McClure received a B.S. degree in chemistry from Grove City College in 1932 and joined Sun as a tank farm gauger in 1933. Four years later he was appointed an analytical chemist at the Marcus Hook refinery laboratory. Since 1944 he has been serving in a liaison capacity between the new solvent refinery unit and the laboratory.

Sun also has advanced Robert J. Crawford from assistant manager to manager of the refinery-accounting department. He succeeds W. Henry Linton, recently appointed assistant to the vice president in charge of manufacturing. Crawford joined Sun Oil Company in 1933.

Appointment of Storrs J. Case as advertising manager of Sun Oil Company also has been announced.

Case, a native of Detroit, has been in advertising or sales promotion in the automotive field for 24 years. Until recently he owned and operated a tire distributorship at Van Nuys, Calif.

In his new position, he will direct all of Sun Oil Company's advertising. Fred S. Cannan will be assistant manager in charge of motor products advertising, and Laurens H. Fritz assistant manager in charge of industrial products advertising.

First Regional Meet 1948 AIChE in New Orleans

The American Institute of Chemical Engineers will hold its first 1948 regional meeting at the Roosevelt Hotel in New Orleans on February 15-19.

E. A. Gastrock, head, engineering and development division at the USDA Southern Regional Research Laboratory in New Orleans, is general chairman. Two symposia are being planned on "High Temperature Refractories" and "Low Temperature Insulation" along with several papers. W. E. Lobo, director of the chemical engineering department, M. W. Kellogg Company, New York, is arranging these two symposia.

Visits will be made to a dozen or more industrial plants in the New Orleans area.

Indiana Standard Helping Veterans With Education

A total of 180 World War II veterans who are employes of Standard Oil Company (Indiana) have been granted leaves of absence by the company to attend

STERLING ON SILVER . .

. . . And Now The Stamp of Quality on A. P. I. RING GASKETS

Sterling MEANS QUALITY

Produced to Sterling's unusually high standard of quality, these "Soft Steel" A.P.I. Ring gaskets are precision made and micromatically inspected for assurance of true radius and close tolerance on pitch diameter. They are laboratory checked to maintain the softness in metals as required in A.P.I. ring gaskets.

All Sterling "Soft Steel" rings are cadmium plated as added protection against corrosion. Individual handling safeguards their sealing surfaces against damage. Sterling "Soft Steel" A.P.I. Ring Gaskets are immediately available in most popular sizes and pressures.

Specify Sterling on your next order of ring gaskets, from your supply store.

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school under provisions of the "GI Bill of Rights." Of the total, 118 are from the manufacturing department, 55 from the sales department, and seven from the general office.

To augment their monthly educational allotments from the government, several veterans attending colleges close to Standard of Indiana refineries or sales field offices have retained part-time positions with the company.

Young and Gray Promoted After Gulf Lawyer Dies

Gulf Oil Corporation and Gulf Refining Company late last month announced appointment of Eldon Young and Archie D. Gray as general attorney and associate general attorney, respectively, with Houston headquarters. The appointments follow the death last month of John E. Green, general attorney.

Young, under the direction of the general counsel, will be in charge of the legal problems of all the Gulf companies, except litigation, throughout the area in which the companies' operations are supervised from offices in Texas (Texas, Louisiana, Mississippi, Arkansas and Mexico). Gray, under the direction of the general counsel, will be in charge of all litigation of the Gulf companies in the same area and the litigation of major importance elsewhere in the U. S., officials explained.

Young was graduated from the University of Texas in 1915 and holds the A.B., L.L.B., and L.L.M. degrees. He practiced law in Mexico and Texas and joined the Gulf law department in 1927, remaining in the New York office until 1937 when he was transferred to Houston.

Gray also is a graduate of the University of Texas law school and has been associated with the Gulf law department since 1935.

John E. Green, Jr., joined the Gulf companies in 1919 and had been general attorney since 1929. He died in Houston November 8 following a fall off a horse which he was riding.

Root Refinery Deal

Pan American Petroleum Corporation, a subsidiary of Standard Oil Company (Indiana), announced last month it had an option to buy 40 percent of the stock of Root Petroleum Company which owns a refinery located at El Dorado, Ark.

The bloc is that held by D. P. Hamilton, president of Root. Pan American has until December 19 to exercise its option.

Service Pins Presented

Daugherty Refinery, a division of L. Sonneborn Sons, Inc., last month honored the 60 percent of its 411 employes who have been associated with the organization for ten or more years. Awards were made commemorating the completion of periods of employment ranging from 10 to 40 years.

Daugherty Refinery was acquired by the Sonneborn organization in 1940.

Demand for International Petroleum Exhibit Space Brings About Expansion

Demand for space to exhibit new equipment at the International Petroleum Exposition to be held in Tulsa, Okla., next May 15-22, has compelled IPE directors to launch another major construction project at the exposition grounds as contractors went to work to build 40 additional booths at the southeastern section of the grounds. As soon as these are completed an additional two acres of open display space along Skelly Drive will be graded, crisscrossed with sidewalks and other exhibit facilities to accommodate those seeking space.

This additional construction brings the total of new work up to $140,000 all of which is ahead of schedule. According to over-all plans the new expansion program adds three acres of additional space with facilities to the north side of the grounds, doubles the capacity of the cafeteria, provides for a four-room hospital and adds almost a mile of new paving and sidewalks. The five big exposition-owned structures, similar in size and style to the Texas Building, are being completely renovated and redecorated. Additional booths have been added to both the Texas and California buildings which front on the new Silver Lane "White Way."

An attractively landscaped park and plaza east of the replica of the historic Drake well is being created. This park, along with a remodeled club house, will be dedicated to the pioneers and "old timers" of the industry.

Special Services

Housing, hotel and room accommodations for delegates, visitors, exhibitors, their associates and employes, is being handled by a special Housing Committee and it is hoped that accommodations will be found for all who attend.

The committee reported recently that every hotel room in Tulsa and nearby cities, all first class apartments, tourist courts and rooms in the better homes have been assigned. Meantime the committee continues to list new accommodations and visitors will know where their rooms are located, just what they'll cost and will have a confirmed reservation.

Special services for those who attend will prove a boon to visitors regardless of their classification in the oil industry. A new and enlarged home for the Nomads has been completed on the exposition grounds. Nomads will also have downtown hotel offices to facilitate their many services for oil personnel from abroad. More than 100 highly specialized committees are already organizing programs to have downtown and exposition headquarters with trained service employes who speak the specialist's language and understand his needs and problems. All such services will be free. Before such visitors or delegates leave home to attend the exposition they'll have detailed data on just where and how they can obtain information from their particular group in the industry.

Directory Promised Early

An official directory of exhibitors will be prepared and will be printed earlier and on more extensive lines than ever before. There will be between 800 and 1000 manufacturing and service corporations listed in the book.

Plans for a conference to perfect ways and means of holding the World Petroleum Congress in conjunction with the 1950 exposition will have an important place on the agenda of the scientific and technical committee at the 1948 show, according to W. B. Way, general manager. A top-flight scientist and former chairman of the IPE scientific and technical committee has already made preliminary investigations abroad and followed through in Washington. The last WPC was held in Paris, June 14-19, 1937. Way also reported Dr. Benjamin B. Weatherby, M.A., B.A., Ph.D., has agreed to serve as chairman of the scientific and technical committee for the 25th anniversary exposition. Weatherby, vice president and director of Amerada Petroleum Corporation, Tulsa, is internationally known for his scientific achievements. No previous IPE Hall of Science will even approach, in scope, the importance or magnitude of the displays, exhibits, features and presentations planned for 1948, officials believe. A complete cross-section of the latest developments in every branch of the petroleum industry will be on parade, by the industry, for the industry, when the 1948 exposition is opened May 15.

Holds Gulf Fellowship

Campbell

John M. Campbell has recently been appointed to the Gulf Oil Corporation fellowship which was established in the School of Chemical Engineering at the University of Oklahoma in July, 1947.

Campbell graduated from Iowa State University in 1943. During the war he was connected with the Du Pont Company in Cleveland, Ohio, and Hanford, Wash. He is now an instructor in the School of Chemical Engineering and is also working toward a Master's Degree in chemical engineering.

Monsanto Promotions

Monsanto Chemical Company personnel changes have been announced by the company as follows:

Dr. James H. Lum, former executive director for Monsanto at Clinton Laboratories, Oak Ridge, Tenn., to Melbourne, Australia to become managing director of Monsanto (Australia) Pty. Ltd., which has plants at Melbourne and Brisbane.

Milton Welhoelter, project engineer in the design section, named assistant manager of the design section of the general engineering department and Robert E. Wright, in the design section since 1945, appointed assistant to the section manager.

William H. Winfield named manager

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of the company's newly established department of research which will devote special attention to general business conditions connected with Monsanto's contemplated expansion program as well as conditions affecting the chemical and allied industries.

Herbert J. Krase of Dayton, Ohio, and Sylvester M. Evans of Everett, Mass., promoted to become asssitant directors of the patent department.

Edmund Greene, formerly of Cambridge, Mass., has been named sales promotion and advertising manager of the Merrimac Division. He joined Monsanto's industrial and public relations department, St. Louis, last June.

Shell Opens New Research Laboratory in Houston

One of the nation's largest and most modernly equipped research laboratories devoted to improvement in methods of discovery and production of oil, was dedicated December 5, in Houston, by Shell Oil Company, Inc. The new laboratory will increase Shell's research activities in this field, President H. S. M. Burns explained, and will centralize functions which were formerly carried on in widely scattered localities.

The three story building provides 40,000 square feet for research on the first two floors, about 8000 more for service and accessory machinery on the third.

The research staff is divided into physical, chemical and geological departments, each of which handles the appropriate phase of every problem the laboratory attempts to solve according to Dr. Harold Gershinowitz, laboratory director.

A large number of Shell officials from points throughout the U. S. attended the Houston dedication.

Thompson Main Speaker

Colonel E. O. Thompson, chairman of the Texas Railroad Commission was the principal speaker. He thought that if World War III should come the oil requirements would be in the nature of 8 million barrels daily and pointed out what an important role research must assume in future operations. Better methods of exploration and better means and methods of recovery were both stressed.

"Research is everywhere pointing the way to better and fuller living and so long as we have free enterprise with the incentive motive, research will go forward and men and women of science will be ever increasingly devoting their efforts to true research," he declared.

Thompson pointed out that in 1915 there were about 100 industrial research laboratories in the U. S., and that in 1946 there were 2500 such laboratories employing 130,000 persons.

"Research has found a friendly climate here in this democracy of ours—perhaps the most congenial in the world," the speaker declared. "Let us be thankful that workers here will be equipped to furnish the optimum of convenience and opportunity to find the truth."

Rust Enters Chemical Process Engineering Field

Entering the chemical process design field, The Rust Engineering Company, Pittsburgh, large engineer-constructor firm, has formed a new process engineering division to extend its design and construction services it offers to

Waterman

Wasselle

Patterson

Bass

the chemical, petroleum, and related industries. The new division will make available process engineering in the petroleum, petro-chemical, organic and fine chemical fields, S. M. Rust, Jr., president, said.

The company thus will be able to undertake entire execution of design, drafting, purchasing and construction of chemical process plants. In addition it will offer consulting service for planning, conducting and supervising research and development programs, and extending the results to commercial practice.

To direct the new program, four process engineers, joined the firm December 1, as follows: Dr. John A. Patterson, Dr. Wendell W. Waterman, Leonard A. Wasselle, and Walter L. Bass.

Dr. Patterson, until recently chief project engineer of Publicker Industries, Inc. also has served with Atlantic Refining Company, Standard Oil Development Company, and DuPont Ammonia Corporation.

Dr. Waterman, formerly associate director of development for Publicker, was for nine prior years associated with the Standard Oil Development Company.

Wasselle was for nine years with Charles Pfizer and Co., Inc., prior to going with Publicker as process engineer.

Bass served for nine years as a process engineer with the Lummus Company and later with Publicker as project engineer.

Willcox Named Manager

T. Norman Willcox has been named manager of the methods and equipment laboratory of the plastics division of the General Electric Chemical Department.

Willcox, section engineer of this laboratory since last February, came to the company in 1936 as a member of the student test course. He joined the plastics division at Pittsfield in 1938. He also served in the Taunton Works as a product engineer for one year.

Willcox received both his bachelor's and master's degree in electrical engineering from the Massachusetts Institute of Technology.

Aluminum Tanks Tested As Steel Substitutes

Tests recently conducted at the Underwriters' Laboratories in Chicago on aluminum fuel oil tanks resulted in the failure to meet the specifications demanded by the industry. Although the results were disappointing, the use of aluminum to relieve the acute shortage of steel appears to be the quickest solution in the fabrication of tanks. Changes in design of aluminum tanks were indicated by results of the tests. The tanks showed distortion at a low hydrostatic pressure and failed at the welds at less than 10 pounds pressure.

Shortly after the Chicago tests, a Midwestern tank manufacturer completed hydrostatic pressure tests of an aluminum tank of his design with failure occurring at 40 pounds pressure. Minimum acceptance pressure is 25 pounds.

Cycling Plant Dedicated At Canfield, Mississippi

Mississippi's first pressure-maintenance and cycling plant was formally dedicated November 21 by Governor F. L. Wright of Mississippi. The Canfield plant is jointly owned by Placid Oil Company, The California Company and Gulf Refining Company. Operating will be The California Company, holders of approximately 75 percent of the stock.

The plant is designed to operate at 1600 psi and will inject gas at 4000 psi.

Jersey Gift Includes Fund For Petroleum Research

Standard Oil Company (New Jersey) has contributed $250,000 to the New York University-Bellevue Medical Center Fund including grants of $100,000 for the construction and initial equipment of laboratories in the Institute of Industrial Medicine; $50,000 for clinical facilities in the University Hospital; $50,000 in support of the general program and $50,000 for the support of the original research relating to the petroleum industry. The last amount will be given in equal annual installments over a period of five years, commencing in 1948.

Chancellor Harry Woodburn Chase of New York University characterized the gift as "one of the largest capital gifts ever made by a business corporation to further medical education in America."

Three Oil Men Named on New Home Safety Group

A coast-to-coast committee of 28 national leaders of business, industry, labor, finance and public affairs has been formed to promote the practice of safety more widely in general daily life, the National Safety Council has announced. Home, school, farm and other accidents that have not been adequately reached

by organized preventive efforts, it was declared.

Chairman is William A. Irvin, former president of the United States Steel Corporation and the group includes three oil industry men as follows: W. S. S. Rodgers, chairman, The Texas Company; Harry C. Wiess, president, Humble Oil & Refining Company; and Walter J. Cummings, director, The Texas Company.

NPC Advises Against Return of Oil Rationing

Government rationing or allocation of petroleum products, in the opinion of the National Petroleum Council, would only cause confusion and make the supply problem worse.

A special committee headed by Harry C. Wiess, president of Humble Oil & Refining Company, gave the report December 1 to the council of which Walter S. Hallanan is chairman.

"Government controls cannot increase the quantity of available supplies," Hallanan reminded. "All they can produce are ration cards with attendant confusion and hardship to consumers."

The committee said it appeared there will be insufficient oil supplies to meet demands in certain areas this winter and claimed such shortages were due in large part to transportation.

It asked the government to return surplus government-owned tankers to service as quickly as possible; limit exports insofar as was possible; encourage conservation of fuels and publicize all efforts which have as their purpose fuel conservation.

Bureau of Mines Says Potential Demand For 1948 Oil To Exceed Refining Capacity

SINCE potential demand for petroleum products has increased more sharply than it has been possible to expand facilities for supplying the oils under existing handicaps and shortages, it is indicated that the industry will continue to operate virtually at capacity in 1948, achieving meanwhile some increase in capacity, without being able to meet fully the requirements of its customers.

Even in 1947 this abnormal situation has existed, in which it has not been possible to satisfy all demand fully. Deficiency arose as early as June, 1947, when refiners of the Midwest were obliged to begin allocating gasoline to their dealers, in proportion to former sales, while the dealers in turn had to allocate supplies among their customers. While Midwestern allocations are still necessary, a similar division of supplies of gasoline and other products among dealers and consumers lately has been initiated in the New England and Middle Atlantic states.

Heretofore, it has been the consensus in the industry that shortages could be confined to local and temporary situations, such as those which have prevailed in the Midwest. But the present East Coast shortage was not foreseen, and it now appears that scarcity of oils may last longer and be more serious than was formerly anticipated. Making it difficult for the industry to establish a balance of supply with demand is a number of factors, including shortages of materials for expansion of facilities on the one hand and the constantly increasing demand on the other. The longer it takes to provide extra capacity, the more extra capacity is needed.

European Aid Drain

To further complicate the problem of balancing supply with demand, the aid to Europe plan now confronts the industry, foreshadowing a likely increase in shipments of U. S. and other oils to Europe and a probable stimulation of exports of American-made oil industry equipment to Europe and to the Middle East and the Caribbean area, which will be the principal sources of oil for Europe.

When all factors are considered, it is indicated that any forecast of the operations of the oil industry of the U. S. in 1948 must be essentially an estimate of supplies that will be available rather than the usual forecast of the probable demand for petroleum products. This is something quite unusual in projecting

Forecast of United States Supply and Demand, All Oils, 1948, and Revised Estimate for 1947

Source: U. S. Bureau of Mines Monthly Petroleum Forecast, Nov., 1947

(THOUSANDS OF BARRELS)

ITEM	1946 (Actual)	1947 (Partly Estimated)	Percent Change '46-'47	1948 (Forecast)	Percent Change '47-'48
PRODUCTION					
All Oils	1,850,700	1,980,300	+ 7.0	2,075,300	+ 4.8
(Daily Average)	5,070	5,425	+ 7.0	5,670	+ 4.8
Crude Oil	1,733,400	1,849,900	+ 6.7	1,936,200	+ 4.7
(Daily Average)	4,749	5,068	+ 6.7	5,290	+ 4.7
Other Oils	117,300	130,400	+11.2	139,100	+ 6.7
(Daily Average)	321	357	+11.2	380	+ 6.7
IMPORTS					
All Oils	135,000	157,000	+16.3	166,800	+ 6.2
(Daily Average)	370	430	+16.3	456	+ 6.2
Crude Oil	86,100	97,700	+13.5	100,800	+ 3.2
(Daily Average)	236	268	+13.5	275	+ 3.2
Other Oils	48,900	59,300	+21.3	66,000	+11.3
(Daily Average)	134	162	+21.3	180	+11.3
NEW SUPPLY					
All Oils	1,985,700	2,137,300	+ 7.6	2,242,100	+ 4.9
(Daily Average)	5,440	5,856	+ 7.6	6,126	+ 4.9
STOCKS CHANGE					
All Oils	+43,500	+4,800	—89.0	+ 8,100	+68.8
(Daily Average)	+119	+14	—89.0	+ 22	+68.8
Crude Oil	+6,900	+4,400	—36.2	+ 500	—88.6
(Daily Average)	+19	+12	—36.2	+ 1	—88.6
Other Oils	+36,600	+400	—98.9	+ 7,600	+18.0
(Daily Average)	+100	+1	—98.9	+ 21	+18.0
DEMAND					
All Oils, Total Demand	1,942,200	2,132,500	+ 9.8	2,234,000	+ 4.8
(Daily Average)	5,321	5,842	+ 9.8	6,104	+ 4.8
Domestic Demand	1,791,000	1,969,500	+10.0	2,085,000	+ 5.9
(Daily Average)	4,907	5,396	+10.0	5,697	+ 5.9
Exports	151,200	163,000	+ 7.8	149,000	— 8.6
(Daily Average)	414	447	+ 7.8	407	— 8.6
Crude Oil Exports	40,700	45,500	+11.8	50,000	+ 9.9
(Daily Average)	111	125	+11.8	137	+ 9.9
Refined Oil Exports	110,500	117,500	+ 6.3	99,000	—15.7
(Daily Average)	303	322	+ 6.3	270	—15.7

ITEM	1946 (Actual)	1947 (Partly Estimated)	Percent Change '46-'47	1948 (Forecast)	Percent Change '47-'48
Motor Fuel, Total Demand	780,200	835,400	+ 7.1	876,000	+ 4.9
(Daily Average)	2,138	2,289	+ 7.1	2,393	+ 4.9
Domestic Demand	734,800	790,000	+ 7.5	840,000	+ 6.3
(Daily Average)	2,013	2,164	+ 7.5	2,295	+ 6.3
Exports	45,400	45,400	.....	36,000	—20.7
(Daily Average)	125	125	.....	98	—20.7
Residual Fuel, Total Demand	486,600	524,000	+ 7.7	529,000	+ 1.0
(Daily Average)	1,333	1,436	+ 7.7	1,445	+ 1.0
Domestic Demand	477,100	513,000	+ 7.5	517,000	+ 0.8
(Daily Average)	1,307	1,405	+ 7.5	1,413	+ 0.8
Exports	9,500	11,000	+15.8	12,000	+ 9.1
(Daily Average)	26	30	+15.8	33	+ 9.1
Distillate Fuel, Total Demand	272,300	320,200	+17.6	346,000	+ 8.1
(Daily Average)	746	877	+17.6	945	+ 8.1
Domestic Demand	243,000	290,000	+19.3	328,000	+13.1
(Daily Average)	666	795	+19.3	896	+13.1
Exports	29,300	30,200	+ 3.1	18,000	—40.4
(Daily Average)	80	83	+ 3.1	49	—40.4
Kerosine, Total Demand	97,700	108,400	+11.0	118,000	+ 8.9
(Daily Average)	268	297	+11.0	322	+ 8.9
Domestic Demand	89,200	101,600	+13.2	110,000	+ 8.9
(Daily Average)	245	277	+13.2	301	+ 8.9
Exports	8,500	7,400	—12.9	8,000	+ 8.1
(Daily Average)	23	20	—12.9	22	+ 8.1
All Other Oils, Total Demand	305,400	344,500	+12.8	365,000	+ 6.0
(Daily Average)	837	944	+12.8	997	+ 6.0
Domestic Demand	246,900	275,500	+11.6	290,000	+ 5.3
(Daily Average)	677	755	+11.6	792	+ 5.3
Exports	58,500	69,000	+17.9	75,000	+ 8.7
(Daily Average)	160	189	+17.9	205	+ 8.7
CRUDE NEEDED					
Runs to Stills	1,730,100	1,842,155	+ 6.5	1,926,990	+ 4.6
(Daily Average)	4,740	5,047	+ 6.5	5,265	+ 4.6
Demand for Domestic Crude	1,727,545	1,845,805	+ 6.8	1,935,774	+ 4.9
(Daily Average)	4,733	5,057	+ 6.8	5,289	+ 4.9

STAINLESS STEEL CASTINGS for heat and corrosion resistant applications are now available from TESCO . . . in addition to carbon steel castings. The heat-resistant stainless steel furnace castings pictured here . . . made to specifications for a Gulf Coast Refinery . . . weigh approximately 900 pounds each, are 7 feet long and 3½ feet wide. These TESCO Stainless Steel Castings will resist oxidation at temperatures up to 2000 degrees . . . are readily welded and machined . . . combine good ductility with high tensile and creep strength at high temperatures. TESCO is ready now to deliver stainless steel castings to your specifications from a pound to a ton . . . assuring you the correct alloy analysis to produce the finest heat and corrosion resistant castings available. So make TESCO your first source for quality steel castings. TESCO is the most experienced foundry on the Gulf Coast. Write for details and quotations.

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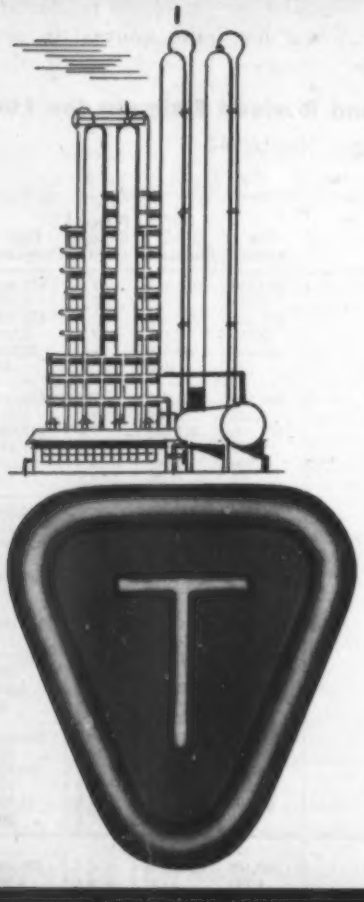

Symbol of Quality Steel Castings

TEXAS ELECTRIC STEEL CASTING CO.

GOOD CASTINGS — CARBON AND ALLOY STEEL TESCO — HOUSTON, TEXAS

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oil industry operations into the future, having become necessary only since shortages began to develop in 1947.

It is upon this extraordinary basis of estimating probable available supply that the Bureau of Mines, U. S. Department of Interior, has offered its forecast of oil industry operations in 1948. In this forecast, the bureau estimates that the total demand for all oils to be supplied by the industry in 1948 will average 6,104,000 barrels daily, an increase of 262,000 barrels per day or 4.8 percent over the 5,842,000 barrels daily indicated for 1947.

In the latter half of 1947, the fulfillment of demand for petroleum products came to be largely a matter of maximum refining capacity, and similarly, in 1948, the amount of demand to be met will be determined mainly by the volumes that can be put through the available refinery facilities. Thus the indicated prospective increase in total demand for oils in 1948 is essentially the same as the indicated possible increase in refinery runs, which the Bureau of Mines estimates at 5,265,000 barrels daily in 1948, an increase of 218,000 barrels per day or 4.6 percent over the estimated average runs of 5,047,000 barrels daily in 1947. An increase for 1948 over 1947 will have to be achieved, if accomplished, in spite of higher than normal operating rates at refineries in 1947. But new refinery capacity is entering the picture to make an increase possible of achievement.

Export Decrease Seen

In its forecast, the Bureau of Mines assumed a decrease in exports to 407,000 barrels daily in 1948 from an indicated 447,000 daily in 1947. On that assumption, it was calculated that the domestic demand which could be fulfilled in 1948 would average 5,697,000 barrels daily, an increase of 301,000 barrels per day or 5.9 percent over the 5,396,000 daily indicated for 1947. Lower exports would leave more for domestic use, it was pointed out, or higher exports would leave less.

It is noteworthy that the forecast for 1948 contemplates a change for the U. S. from the traditional role of net exporter of oil to that of net importer, a change heretofore reflected in monthly but not annual statistics. Against the estimated 407,000 barrels per day of exports in 1948, the U. S. would import 456,000 barrels daily, according to the forecast, a net importation of 49,000 barrels per day. In 1947 U. S. exports of 447,000 barrels daily will exceed imports of 430,000 daily by 17,000 barrels, it is indicated.

In discussing its forecast for 1948, which is embodied in its Monthly Petroleum Forecast for November, 1947, the Bureau of Mines explained that the forecast was, of necessity, "based on the probable supply of oil products as determined by available refinery capacity and the limits of existing transportation facilities to move crude oil to refineries and to distribute products.

"The primary controlling factors," stated the bureau, "is the estimate of crude runs to stills, and it is believed that this estimate is reasonably optimistic after allowing for necessary refinery shutdowns for repairs or emergencies.

"Potential demand for oil products should be substantially greater than the forecast of demand that might be supplied under existing conditions," the agency pointed out.

"Imports can be projected only on the basis of present trends and policies," the explanation continued. "Imports of refined products (chiefly residual fuel oil) supplement the limited output of our own refineries."

ASME Hears of Industry's March Toward New and Better Products at Annual Meet

"Production for Peace" was the subject of the 68th annual meeting of The American Society of Mechanical Engineers held in Atlantic City December 1 to 5. The onward march of industry toward new and better products, more efficient methods, and greater economy of production, and the engineers' part in that progress, keynoted many of the 80 technical sessions, with some 240 authors and speakers.

A special research committee on the properties of gases and gas mixtures has been set up by the engineering society. To supply fuller knowledge, allowing for revision and growth in reference data as new facts become known, will be the purpose of the ASME committee.

Petroleum refining, which has been called the largest chemical industry in the world, has an extensive need for gas properties data. Survey of the industry indicated a need for high-pressure gas data for data on process gases.

Priority of interest, according to the ASME survey, was indicated for groups of gases, in the order named as follows: basic gases, hydrocarbons, gas mixtures, refrigerants and process gases.

Major interest centered in thermodynamic properties of constant pressure specific heat, constant volume specific heat, internal energy, enthalpy, entropy, and compressibility; physical properties of viscosity, thermal conductivity, and emissivity.

In the hydrocarbon group, of primary importance to the petroleum industry, additional compressibility and vapor phase data are needed especially for such specific compounds as pure and oxygenated hydrocarbons, through the C^4 range.

In gas mixtures, such as oxygen-nitrogen, nitrogen-methane, carbon monoxide-water vapor, and others, a serious lack of information exists. Many operations in petroleum refining, gas separations, and combustion, involve great variety of mixtures.

Establishment of a synthetic liquid fuels industry in the United States, to augment present petroleum resources, should not be delayed "until an emergency period," W. C. Schroeder of the Bureau of Mines told the fuels session.

While there is a known reserve of 21 billion barrels plus a small net import "it is evident that domestic reserves must soon be increased or new sources of supply found." The picture is "disquieting" since with mounting consumption, new oil pools for eight or nine years have failed to match it.

As for the importation of foreign oil, Schroeder continued, the proportion available to the United States tends to diminish as foreign countries are turning more to its use.

Combustion engineers are meeting the challenge for advancing costs of standard fuels by developing new types of furnaces which will burn low-cost substitute fuels in industrial processes. These industrial by-products, formerly regarded as necessary waste, may be used to supplement and even to replace standard fuels such as coal, oil and gas, thereby saving "fuel dollars" for the manufacturer, Otto de Lorenzi, director of education of the Combustion Engineering Company, New York, said in presenting his paper, "Furnaces for By-Products Fuel."

Technical descriptions and drawings of furnace designs now available for burning these erstwhile "waste" materials were supplied by de Lorenzi.

Big Staff Needed if Oil Controls Are Promulgated

Oil control, if voted by the Congress, would require a staff of between 500 and 1000 employes and involve a government cost of from $2 to $4 million yearly.

These are the fiugres which Washington last month thought would be passed on by the Department of the Interior to the Budget Bureau, as Acting Secretary Chapman gave his views on price control to a House committee. Interior meanwhile says it needs $100,000 to be used in developing public support for the conservation program which President Truman had in mind.

Chapman also is seeking $250,000 with which the Bureau of Mines would compile current inventory and other oil statistics which he says are essential both to the success of any voluntary measures designed to meet the tight oil situation and "to the speedy institution and the effectiveness of stringent controls, if their use should be needed."

Houdry School Plan

An education-refund plan has been made available to all employes of Houdry Process Corporation and provides for reimbursements to all participating employes who undertake approved study courses for their own advancement.

Refinery and Plant Construction Notes

(Also see compilation of construction projects, pages 156-157)

California Standard To Build Utah Refinery

Standard Oil Company of California has announced plans for the construction of a 25,000-barrel-daily refinery unit in the Salt Lake City area, actual construction to start in the early spring. Completion of the $5 million plant is anticipated in about 12 months. This will coincide with the completion of the pipe line being constructed by The California Company from the Rangely field, Colo.

This initial refinery installation will include crude oil distillation and other facilities for handling and processing of crude into a limited number of products, including gasoline, heating oils and fuel oils. The type and amount of supplementary facilities have not been decided upon, officials declared, but this first unit is designed to fit in with much larger complementary facilities. The crude distillation unit will be built by the Bechtel Corporation.

Northwestern to Enlarge Comet Refining Plant

Plans for modernization and enlargement of Comet Refining Company's plant at New Brighton, Minn., are being studied by officials of Northwestern Refining Company, Saint Paul Park, Minn., which recently purchased the property.

The plant, a 3000-barrel skimming unit, primarily produces heavy fuel oil and asphalt. An increase in capacity of the Comet refinery is anticipated but the first improvement will be modernization and largement of the sour crude treating facilities, it was explained.

Richard G. Gray, son of the former owner of Comet, will continue in an executive capacity with Northwestern. Stanley G. Gray, former owner of Comet, plans to retire from the oil business.

Co-op Adding Unit

Construction of a 5000-barrel-daily catalytic reforming and catalytic desulfurization unit at the National Co-operative Refinery Association's McPherson, Kan., plant was announced last month by E. J. Mary, general manager. Initial operation of the unit is scheduled within the next two months.

The unit is licensed by the Perco division of Phillips Petroleum Company and is being constructed by Koch Engineering Company of Wichita, Kan.

Utah Plant Expansion

Expansion of the refining facilities purchased from the federal government has been announced by Utah Oil Refining Company. The refinery, built at Salt Lake City during the war, will have a daily charge capacity of 25,000 barrels of crude oil. Cost of the expansion program will approximate $2 million.

Plant at Levelland, Texas

Stanolind Oil & Gas Company will build a natural gasoline plant in the Levelland field, Hockley County, Texas. Other producers in the field will be given the opportunity to participate in the plant ownership.

Pressure maintenance is being investigated, and if it should be found desirable the plant will be equipped with facilities for returning high-pressure gas to the leases for reinjection, officials said.

Ashland Buys WAA Plant Located at Catlettsburg

A government-owned refinery near Catlettsburg, Ky., has been sold to its wartime operator, Ashland Oil and Refining Company, of Ashland, Ky., for $2,150,000. The sales price covers 110 acres of land, production equipment, and numerous buildings..

The plant includes three processing units and numerous accessory facilities. The units are a Thermofor catalytic cracking and treating unit and concentration or feed preparation unit, a Universal Oil Products H. F. alkylation unit, and a Universal Oil Products butane isomerization unit.

New Gasoline Plant Building at Carthage

Work has begun on a new gasoline plant five miles east of Carthage, Texas, in Panola County, for Glassell and Glassell of Shreveport, La. The gas processed in the plant will be purchased by Tennessee Gas & Transmission Company and transmitted through a feed line to their main line.

Price of the stripped gas has been reported on a scale of six and one-half to eight cents per 1000 cubic feet.

Large Texaco Terminal

One of its largest terminals in the south, to be located adjacent to the company's West Dallas refinery, will be erected by The Texas Company with work scheduled to start immediately. The terminal will have a capacity of approximately 18 million gallons and also large warehouse and garage facilities.

Petrochemical Plant

British Petroleum Chemicals, Ltd., is planning construction of a petrochemical plant at Grangemouth, Scotland, with plans awaiting approval of the British government. The $22 million plant will produce derivatives of the lower olefins.

British Petroleum Chemicals, Ltd., is jointly owned by the Anglo-Iranian Oil Company and Distillers Company, Ltd.

Anglo-Iranian also is planning large scale expansion of its refinery at Grangemouth.

Refinery for England

Early November heard Petrocarbon, Ltd.'s announcement of plans to build a new 20,000-barrel-daily refinery near Manchester, England. Moore & Charlton, consulting engineers of New York, who made the announcement, said completion was scheduled for early 1949. The site will be adjacent to a plant being built by Petrochemicals, Ltd., for manufacture of aromatics, ethylene and other petro-chemicals.

The projected refinery will include a 5000-barrel-daily catalytic cracker, reforming units and facilities for lube manufacturing.

Catalytic Construction Readying Petrol's Plant

T. Ellwood Webster, President of Catalytic Construction Company, subsidiary of Houdry Process Corporation, has announced that work has begun on the Texas City Refinery of Petrol Terminal Corporation which the latter recently purchased from War Assets Administration. The plant includes a 6-case Houdry fixed-bed catalytic cracker and work will embrace inspection, repair, and testing of the catalytic unit as well as the alkylation plant and gas plant.

The unit has been down for two years and modernization will be included in the maintenance work by Catalytic Construction engineers. It is expected that the plant will be back on stream in less than 90 days.

Catalytic Construction has also begun maintenance work at Socony-Vacuum Oil Company, Inc's Buffalo refinery, changing catalyst and making complete repairs to Houdry fixed-bed cases. While the work is being done on a one-shift per-day basis the job will be completed in less than six weeks.

In Washington WAA predicted that Petrol Terminals Corporation, which is headed by E. M. Callis, would buy the surplus toluene plant at Beaumont, Texas (operated by Magnolia Petroleum Company during the war) for $360,000, for the purpose of moving the equipment needed for crude topping to Texas City. Callis expected to have 30,000 barrels daily crude capacity at Texas City and 4000 barrels daily production of aviation gas on stream by January 1.

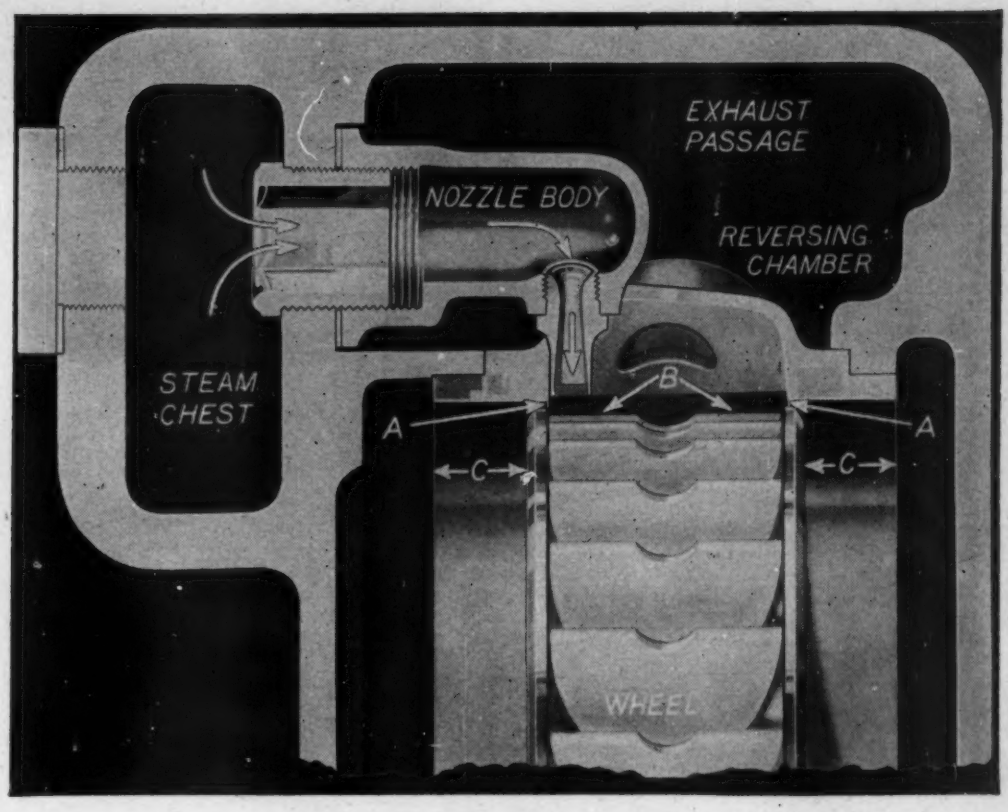
Russian Bound Equipment Diverted to U. S. Industry

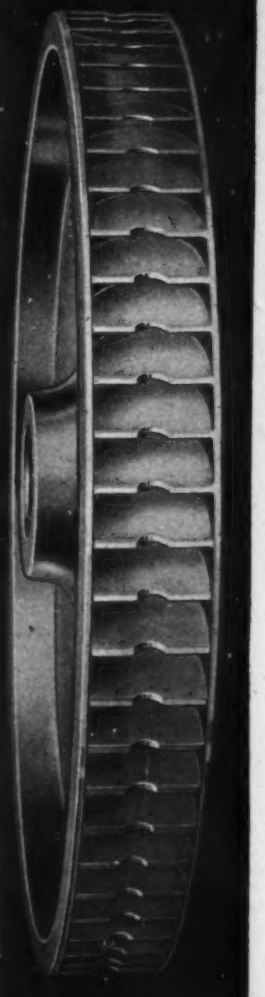
Engineers are now inventorying and appraising approximately $7 million worth of gasoline cracking plant equipment originally intended for shipment to Russia under terms of lend-lease but having failed to be shipped before the deadline date, remained in this country and will be turned over to WAA for disposal as surplus.

While no complete plant is involved which would process crude completely into gasoline, whole units for addition to existing plants to expand their capacity will be made available to the refining industry through the WAA, George Truesdale of the Lend-Lease Division of the State Department, announced last month. The thousands of pieces of equipment will be turned over to the WAA within a month.

The equipment was originally scheduled for use in the Soviet Union to expand the capacity of existing refineries in Russia. As it was not shipped before the deadline date of December 31, 1946, it has remained in this country.

TERRY





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SCIENCE *and* TECHNOLOGY ABSTRACTS

The abstracts here presented are selected from the current literature of science and technology to afford reference to fundamental information not easily available to all readers. Abstracts of articles appearing in readily obtainable trade journals are not included.

Photostat copies of original articles will be supplied at cost by The Leslie Laboratories. Complete or limited bibliographies covering special topics by title, by abstracts, or in complete manuscript, will also be prepared and furnished at reasonable cost by the Laboratories.

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Fundamental Physical and Chemical Data

Heat Capacities of Gaseous Oxygen, Isobutane, and 1-Butene from —30° to 90° C. P. F. WACKER, R. K. CHENEY AND R. B. SCOTT, *J. Research Natl. Bur. Standards* **38** (1947) pp. 651-9.

Using an improved calorimeter, measurements of the heat capacities of isobutane and 1-butene were made in order to make data available for use in combination with spectroscopic data for the calculation of thermodynamic functions over a range of temperatures. Heat capacity measurements of isobutane were made at —30, 0, +40, and 90° C. A few measurements were made on 1-butene at 40° and 90° C. Observations were made at two or more pressures, except for 1-butene at 90° C. At —30° C. the measurements on isobutane were made at low pressures to avoid condensation of the gas as it passed through the calorimeter. The data are presented in tabular form.

The Thermodynamic Properties and Molecular Structure of Cyclohexane, Methylcyclohexane, Ethylcyclohexane and the Seven Dimethylcyclohexanes. CHARLES W. BECKETT, KENNETH S. PITZER AND RALPH SPITZER, *Jour. Am. Chem. Soc.* **69** (1947) pp. 2488-95.

The paper is a report on continued studies of the structural and thermodynamic properties of hydrocarbons. Data for clyclohexane from electron diffraction, Raman and infra-red spectra, and thermodynamic measurements were analyzed in various ways to give an assignment of vibration frequencies and calculated values of the heat capacity, entropy and heat and free energy functions for the hydrocarbons in the ideal gas state from 298.16 to 1500° K. It is concluded that at room temperature cyclohexane is chiefly in the chair form of symmetry D_{3d}. However, the heat capacity of gaseous cyclohexane shows a contribution from conversion to the boat form which is approximately 5.6 kcal./mole higher in energy.

The possibilities of geometrical tautomerism in the methyl and dimethylcyclohexanes are discussed in detail. It is shown that energy differences between these tautomers are related to the strain or steric energies in *n*-paraffins, and that there is good agreement between the corresponding energy quantity obtained from data on each series of compounds. The heats of isomerization confirm the assignment of the configuration *cis*-1, 3-dimethylcyclohexane to the isomer boiling at 120.1° C. and *trans*-1,3 to the isomer boiling at 124.5° C. Thermodynamic functions mentioned were calculated also for methyl, ethyl and for seven dimethylcyclohexanes in the ideal gas state from 298.16 to 1500° K. Data are presented in considerable detail in tabular form.

The Thermodynamics and Molecular Structure of Cyclopentane. JOHN E. KILPATRICK, KENNETH S. PITZER AND RALPH SPITZER, *Jour. Am. Chem. Soc.* **69** (1947) pp. 2483-8.

The published data from electron diffraction, Raman and infra-red spectra, entropy and specific heat measurements were combined and studied in order to develop as complete a picture of the structure of cyclopentane as possible. It was found that the puckering of the ring is not of a definte type, but that the angle of maximum puckering rotates around the ring. Other properties are consistent with this conclusion. Vibration frequencies are defined and tables of the thermodynamic functions are given for the range 298.16 to 1500° K.

The Enthalpy, Entropy and Specific Heat of Liquid p-Xylene from 0 to 300°. The Heat of Fusion. R. J. CORRUCCINI AND D. C. GINNINGS, *Jour. Am. Chem. Soc.* **69** (1947) pp. 2291-4.

The author has found it possible to obtain fairly accurate values for the enthalpy referred to 0° C. of saturated liquids at temperatures approaching the critical by using an ice calorimeter. Only small samples of a few ml. are required. Experimental technique is described in some detail. The enthalpy, entropy and specific heat of saturated liquid *p*-xylene were determined in the range, 0 to 300° C. The heat of fusion of *p*-xylene was determined. The data are presented in tabular and graphical form.

Experimental Vapor Heat Capacities and Heats of Vaporization of n-Hexane and 2,2-Dimethylbutane. GUY WADDINGTON AND DONALD R. DOUSLIN, *Jour. Am. Chem. Soc.* **69** (1947) pp. 2275-9.

The current thermodynamic program of the Petroleum and Natural Gas Division of the Bureau of Mines includes the measurement of the heat capacities of hydrocarbon vapors. Incidental to this, values of the heats of vaporization are obtained. The apparatus used in the work has been described in a recent publication. The present paper gives the results of measurements of the vapor heat capacities and the heats of vaporization of *n*-hexane and 2,2-dimethylbutane. The procedure used in the work is described, and the data obtained are presented in tabular form. The heat capacity measurements were made at several pressures in the temperature range 33 to 470° K. Equations are given to express the capacities in terms of the absolute temperature. Equations are also given that permit calculation of the latent heat as a function of temperature. Second virial coefficients which represent the low temperature low pressure behavior of the vapors were obtained from thermal measurements.

Pressure - Volume - Temperature Relations of Benzene. EDWARD J. BORNOWSKI, ERWIN H. AMICK, JR. AND ARTHUR NORMAN HIXSON, *Ind. and Eng. Chem.* **39** (1947) pp. 1348-52.

The authors note that although considerable information has been published on the pressure-volume-temperature relations of the paraffins and to a lesser extent on the olefins, few data have been published on the aromatics, or on binary solutions either in whole or in part aromatic. The vapor pressure of benzene was determined from 130° C. to the critical point. Critical constants were found to be as follows: Critical temperature 289.5° C., critical pressure 36,985 mm., and critical volume 3.36 ml. per gram. The pressure-volume-temperature relations were determined and the compressibility factors calculated for temperatures ranging from 240° to 355° C., and for pressures ranging from 19,000 to 50,000 mm. It is shown that the generalized compressibility factor charts for hydrocarbons, which are based on the data for

Chemicals Wanted

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2,6-Dimethylnonane
1,4-Dimercaptobutane
Bis(3-Methyl-3-hexene)-4-sulfide
Bis(3-Methyl-3-heptene)-5-sulfide
3,5-Quinazoline
α-Methylcyclopentanone
5-Chloropentene-1
6-Chlorohexene-1
7-Chloroheptene-1
8-Chlorooctene-1
Thapsic acid
Brassylic acid
Eicosyl alcohol
Brazan
Xanthene
1,2-Dichloroisobutane
1,3-Dichloroisobutane
2-Methylheptadiene-1,5
Decadiene-3,7
3-Ethyloctadiene-1,5
3,4-Diethylhexadiene-1,5

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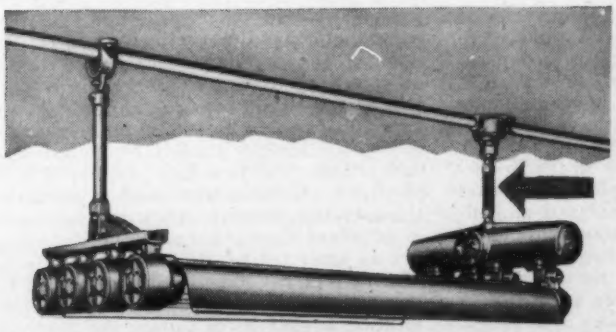

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paraffin hydrocarbons, can be safely used for estimating the pressure-volume-temperature relations of benzene. The apparatus and procedure used in the work are described and the data are presented in considerable detail in tabular and graphical form. A bibliography of 22 references is included.

Correlation of Vapor-Liquid Equilibria Data for Nonideal Ternary Systems. EDWARD G. SCHEIBEL AND DANIEL FRIEDLAND, *Ind. and Eng. Chem.* **39** (1947) pp. 1329-34.

The field of azeotropic distillation and extractive distillation to separate close boiling compounds has been receiving increasing attention in recent years. A few large scale applications have been made. The separations effected in these processes depend on the deviation of vapor pressure from ideality as defined by Raoult's law. There has been no general method for correlating or predicting the vapor-liquid relations of a ternary system from binary equilibrium data. A system is proposed for correlating and predicting the vapor-liquid equilibrium data for a ternary system from the data for the three binary systems. The method is based on an interpolation of the activity coefficients for the ternary system. The method has been found to predict bubble points to an accuracy of 1° C., and vapor compositions with an average deviation of approximately 2 percent of less. The agreement appears to be almost within the limits of reliability of the original ternary data with which comparisons were made. A bibliography of 38 references is included.

Raman Spectra of Hydrocarbons. M. R. FENSKE, W. G. BRAUN, R. V. WIEGAND, DOROTHY QUIGGLE, R. H. MCCORMICK AND D. H. BANK, *Ind. and Eng. Chem., Anal. Ed.* **19** (1947) pp. 700-65.

The Raman spectra of 172 pure hydrocarbons are given both as reproductions of the original records obtained from a recording spectrograph and in tabular form as scattering coefficients and depolarization factors. Data are given for 76 paraffins, 32 naphthenes, 29 olefins, 3 diolefins, 30 aromatics, and 2 other hydrocarbons. Spectra as given are comparable directly because a uniform intensity scale was used. The spectrograph employed records the Raman spectrum as a graph with coordinates linear as to wave length and intensity. Raman spectra are used for the qualitative and quantitative analysis of hydrocarbon solutions. A few solutions containing as many as nine components have been analyzed successfully. In general, Raman spectroscopy is best used in hydrocarbon analyses as a complement rather than as a substitute for infrared and ultraviolet spectra. The instruments used and the analytical procedure and technique are described and discussed. The analyses of known mixtures are given. A bibliography of 27 references is included.

Distribution in Hydrocarbon-Solvent Systems. T. G. HUNTER AND T. BROWN, *Ind. and Eng. Chem.* **39** (1947) pp. 1343-5.

During a study of the separation by solvent extraction of aromatic naphthene and paraffin hydrocarbon solutions, several ternary hydrocarbon-solvent systems were studied. Liquid-liquid equilibria in ternary systems composed of aniline and two hydrocarbons were determined. Data are given for the systems: aniline-cetane-benzene, aniline-cetane-cyclohexane, aniline-heptane-cyclohexane, and aniline-cetane-*n*-heptane. The authors point out that to the best of their knowledge only three ternary hydrocarbon-solvent systems have been studied and equilibrium data reported on them.

Batch Distillation Nomograph for Binary or Multicomponent Mixtures. WALTER H. STANTON, *Ind. and Eng. Chem.* **39** (1947) pp. 1042-3.

A nomograph of the Rayleigh equation for dealing with multicomponent solutions in simple batch distillation is presented. This nomograph is applicable either to binary solutions or multicomponent solutions. It deals in total pounds (or moles) rather than in pound fractions (or mole fractions). The use is limited to the range in which α, in the Rayleigh equation, remains fairly constant. When α is not constant the authors suggest other procedures for use.

Nomograph for Distillation of Low-Boiling Hydrocarbons. FRED M. NELSEN, FRANCIS R. BROOKS AND VICTOR ZAHN, *Ind. and Eng. Chem., Anal. Ed.* **19** (1947) pp. 814-15.

Nomographs are given for converting distillation temperatures and pressures to more familiar boiling points at 760-mm pressure. To cover the full temperature range encountered in distillation of low-boiling hydrocarbons, it has been found most convenient to construct two nomographs, one covering the distillation range +21° to —85° C. and the other —80° to —175° C. The manner in which the nomographs were constructed is described.

Chemical Compositions and Reactions

Kinetics and Mechanism of the Thermal Decomposition of n-Heptane. W. G. APPLEBY, W. H. AVERY AND W. K. MEERBOTT, *Jour. Am. Chem. Soc.* **69** (1947) pp. 2279-85.

Thermal decomposition of *n*-heptane was studied by a flow method in the temperature range 550 to 630° C. and at pressures of 1 to 8.7 atmospheres. Complete analysis of the reaction product was made by use of a method of multi-isothermal distillation for determination of the gaseous products, combined with olefin analysis of the liquid product. In agreement with the predictions of the Rice theory and experimental observations of Dintses, only traces were found of saturated products having more carbon atoms than propane. The important simplifications in the analytical requirements for kinetic studies of higher hydrocarbons that this fact permits is emphasized. Composition and concentrations of the decomposition products are in excellent agreement with the Rice theory as amended by Rice and Kossiakoff. The rate of decomposition was found to be of modified first order type where inhibition by products causes the first order rate constant to decrease with increase of conversion. The activation energy, calculated from the temperature variation of the first order rate, in 65,000 cal.

Epsilon Isomer of 1,2,3,4,5,6-Hexachlorocyclohexane. K. C. KAUER, R. B. DUVALL AND F. N. ALQUIST, *Ind. and Eng. Chem.* **39** (1947) pp. 1335-8.

The addition of chlorine to benzene results in the formation of several isomeric forms of 1, 2, 3, 4, 5, 6-hexachlorocyclohexane. A recent article on the subject described four isomers, the alpha, beta, gamma, and delta, of a possible sixteen isomers that are proposed. In the present paper the authors are concerned with a report on a study of the isolation of the epsilon isomer from the chlorination process product. The physical and chemical properties of the new isomer are given, as well as infrared absorption spectra, X-ray powder diffraction patterns, and solubility data. It was concluded that the epsilon isomer has been isolated. A bibliography of 17 references is included.

Kinetics of Carbiding and Hydrocarbon Synthesis with Cobalt Fischer-Tropsch Catalysts. SOL WELLER, *Jour. Am. Chem. Soc.* **69** (1947) pp. 2432-6.

In the Fischer-Tropsch process for the production of hydrocarbons from carbon monoxide and hydrogen cobalt or iron catalysts are commonly used at temperatures of 170 to 300° C. Straight-chain hydrocarbons are produced almost exclusively, the relative proportions of saturated and unsaturated components varying with the catalyst and conditions. Rates of carbiding, of hydrogenation of carbide, and of hydrocarbon synthesis from carbon monoxide-hydrogen mixtures were measured for a cobalt-thoria-kieselguhr catalyst (100:18:100) by use of a circulating gas system. The initial carbiding rate is comparable to the synthesis rate, but the steady rate of bulk carbiding is about one-tenth as large. The apparent activation energies for bulk carbiding and for the synthesis are 31 and 26 kcal./mole, respectively. Within the pressure range 8 to 45 cm. the pressure dependence of the carbiding rate can be expressed as Rate $= kp^n$, where n is 0.20-0.26. Some of the implications of the experimental results for the mechanism of the carbiding and synthesis are discussed. The apparatus is described is some detail and the experimental data are presented in tabular and graphical form.

Manufacture: Processes and Plant

Solvent Separation of Hydrocarbon Mixtures by Vapor-Liquid Extraction. M. R. FENSKE, C. S. CARLSON AND D. QUIGGLE, *Ind. and Eng. Chem.* **39** (1947) pp. 1322-8.

The purpose of the work was to obtain equilibrium data sufficient to prevent the establishment of a procedure for evaluating methods of separating hydrocarbon mixtures by vapor-liquid extraction. Vapor-liquid equilibrium data at one atmosphere pressure were obtained for the system methylcyclohexane-toluene containing various proportions of aniline to alter the volatility of this pair of hydrocarbons. These data illustrate various features of vapor-liquid extraction or so-called extractive distillation. The term "gamma" is used to define the volatilities that exist in the presence of solvents in order to prevent confusion with ordinary distillation processes. A compact, rugged, vapor-liquid equilibrium apparatus that is suitable for mixtures that give liquid-phase separation on cooling is described. Simultaneous determination of equilibrium temperatures and vapor-liquid equilibria are possible with this apparatus. The relative volatility of methylcyclohexane-toluene solution is increased by the presence of aniline in the liquid phase, the extent of

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Duplex Tubing is made up of two different metals, each designed to resist the particular corrosive element involved on the inside and outside of the tubing. It is being successfully used for ammonia refrigeration and the manufacture of ammonia and its compounds. It is also being tried out for the manufacture of dyes, soaps, drugs, resins, paints, varnishes, food processing and many other products. Often the use of copper-base alloys results in improved heat transfer properties and higher efficiency. Corrosion-resisting copper-base alloys are finding ever greater uses in the manufacture of equipment, piping, valves, and controls.

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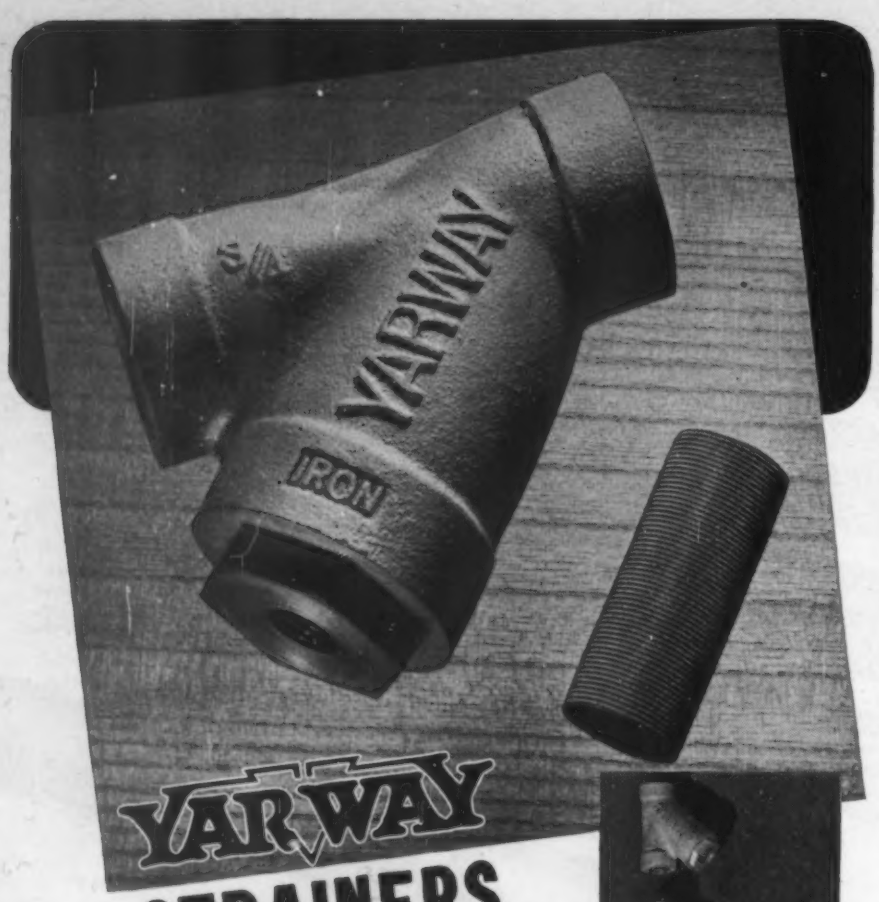

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the increase being dependent on the concentration of aniline in the liquid. Once the equilibrium data are available, design calculations for distillation equipment can be made along essentially conventional lines, using the usual heat and material balances. A bibliography of 20 references is included.

Hydrogenation of Catalytically Cracked Naphthas for Production of Aviation Gasolines. ALEXIS VOORHIES, JR., W. M. SMITH AND C. E. HEMMINGER, *Ind. Eng. Chem.* **39** (1947) pp. 1104-7.

Hydrogenation of the olefinic constituents of catalytically cracked aviation gasolines to the corresponding paraffins without hydrogenation of aromatic constituents to naphthenes results in a gasoline of improved stability and lean-mixture octane number after addition of lead, with no sacrifice in rich-mixture rating. Commercial operation of the hydrogenation process for the improvement of fluid-catalyst-cracked aviation gasolines has been successfully effected both at 150 and at 3000 pounds per square inch pressure using rugged sulfur-resistant catalysts of satisfactory activity and life. The laboratory development of the procedure is first described and the large scale application is briefly considered. Data obtained in the work are presented in some detail in tabular and graphical form.

Esso's Fluid Catalyst Modifies Fischer-Tropsch. JAMES A. LEE, *Managing Editor, Chem. Eng., Chem. Eng.* **54** (1947) No. 10, pp. 105-7.

Conversion of natural gas and coal into such liquid fuels as gasoline and diesel oil has become an important matter in current research and development programs of the petroleum industry. This is not because of fear of a future oil shortage, but has been brought about by the high cost of locating and producing oil. A pilot plant is being built at Library, Pa., near Pittsburgh, to be completed in early 1948. Standard Oil Development Company now has eight pilot plants in operation. In these plants hydrocarbons are synthesized from natural gas by fluid catalyst techniques. The present article deals with pilot plants at Baton Rouge, La., in which hydrocarbons are synthesized from natural gas by fluid catalyst techniques. The limitations of the German type of plant with its fixed type of bed are described. The number of reactors required for a plant producing 10,000 bbl. a day of synthetic gasoline has been reduced to six, or one-twentieth the number required by the German fixed-bed design. The cooling surface has been reduced from nearly 6 million sq. ft. in the original Fischer-Tropsch process to 35,000 sq. ft., or less than one-half of one percent of that needed by the German plant. Types of catalysts have been improved. Those used by the Germans gave highly paraffinic products, so that the gasoline obtained was low in octane number. The catalyst considered for application in this country is of different type and gives mostly olefinic materials, so that the gasoline is of relatively high octane number. A cooling surface is provided in the fluidized bed to remove the high heat of reaction which amounts to 1.7 Btu per bbl. of liquid products. The hydrocarbons and water formed are condensed from the vapor stream. Residual gas is partly recycled to the reactor. The light hydrocarbons are recovered. Residual gas can be used for heat or

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***Chemical Compositions**

	Nicloy 3½	Nicloy 5	Nicloy 9
% Carbon	.20 Max.	14 Max.	.12 Max.
% Manganese	.39 - .60	.30 - .60	.30 - .60
% Phosphorus	.040 Max.	.040 Max.	.030 Max.
% Sulphur	.040 Max.	.040 Max.	.030 Max.
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% Nickel	3.25 - 3.75	4.75 - 5.25	8.00 - 10.00

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power requirements or used to make more synthesis gas. Entrained catalyst is recovered in cyclone separators. Gasoline, after mild treatment, has an ASTM unleaded octane number of about 80 and a research unleaded number of about 93. Some diesel fuel, alcohols, and acids are produced. The oxygenated compounds are largely lower alcohols, methyl through amyl, and the lower fatty acids, acetic through butyric. Some acetaldehyde and acetone are formed. It is estimated that a plant large enough to produce 8,300 bbl. per day of gasoline will cost approximately $35 million exclusive of the investment for the recovery of oxygenated compounds. With natural gas at five cents per 1000 cu. ft. gasoline can be made at a price competitive with its production from petroleum today.

Diisopropyl from Isobutane-Ethylene Alkylation in Presence of BF_3-H_2O-HF Catalyst Systems. W. Nelson Axe and Walter A. Schulze, *Ind. & Eng. Chem.* **39** (1947) pp. 1273-9.

In recent years diisopropyl (2, 3-dimethylbutane) has become important as a blending agent in aviation gasoline because of its high antiknock rating, especially under supercharged rich-mixture conditions. The study was undertaken of its production by the direct alkylation of isobutane with ethylene in the presence of BF_3 — H_2O — HF catalyst complex. The catalyst system resembles the $AlCl_3$— hydrocarbon complex in its capacity to promote the reaction of isobutane and ethylene at temperatures of 100-135° F. and at pressures usually not higher than 200-250 pounds per square inch to yield alkylate containing from 50 to 70 volume percent of diisopropyl. Since boron fluoride is more expensive than aluminum chloride, experimental work was directed mainly towards high alkylate yields with a minimum consumption of boron fluoride. Catalyst prepared by saturating 40 to 50 percent aqueous hydrogen fluoride with boron fluoride is capable of producing only about 8 to 10 volumes of alkylate per volume of catalyst at full activity. However, if the catalyst is continuously activated with boron fluoride, as much as 88 volumes can be produced before signs of failure are evident. In addition to activation, further extension of catalyst life can be effected by fortification of partially spent catalyst with anhydrous hydrogen fluoride, whereupon the system is one again responsive to boron fluoride activation. Results are reported showing the production of 195 volumes of alkylate per volume of catalyst. Additional data are presented to show the effect of hydrogen fluoride concentration on the catalyst activity, external isobutane-ethylene mole ratio and temperature on yield and quality of alkylate, and hydrocarbon impurities in the ethylene on alkylate quality. A bibliography of six references is included.

Chemical Engineering Materials of Construction Review. Editorial Staff, *Ind. & Eng. Chem.* **39** (1947) pp. 1193-1264.

This is the first annual review of chemical engineering materials of construction presented by Industrial and Engineering Chemistry. The large volume of technical information being published today practically demands a periodic recapitulation of progress if the technical man is to keep abreast of developments vitally important to his

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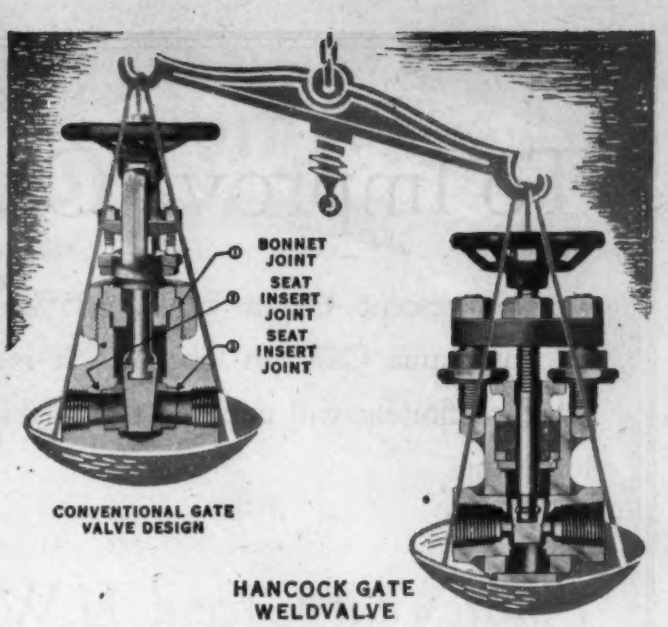

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work. Literature of construction materials is covered approximately from the start of World War II. Subsequent reviews will extend it to form a continuous record from that time. Among the subjects covered are aluminum alloys; carbon and graphite; wrought copper and copper-base alloys; elastomers; fibers; glass; iron, mild steels, and low alloy steels; lead, tin, zinc, and their alloys; nickel and high-nickel alloys; plastics; refractories; hard rubber; stainless steel and other ferrous alloys; chemical stoneware; wood; and finally a codification of materials. Each section includes a bibliography of technical articles dealing with the particular subject.

Products: Properties, Utilization and Analysis

Analysis of a Natural Gas by Volumetric Chemical Methods and by Mass Spectrometer. MARTIN SHEPHERD, *Ind. & Eng. Chem., Analytical Ed.* **19** (1947) pp. 635-40.

The task of standardizing the methods of analysis of gaseous fuels has been undertaken by Subcommittee VII of Committee D-3 of the American Society for Testing Materials. In this program a standard sample of natural gas was analyzed by 50 laboratories in cooperation with the committee. Special chemical methods were used by 30 laboratories. The other 20 employed the mass spectrometer. The results of the work are compared in the paper, and are presented in the form of frequency-distribution plots.

A New Method for the Presentation of Test Results on the Four-Ball Machine. W. DAVEY, *Jour. Inst. Petroleum* **33** (1947) pp. 574-7.

The author presents what he believes to be a more satisfactory method of presentation of the results obtained on the four-ball machine. It is suggested that the wear-load relation be expressed as a wear-load ratio, that the maximum load tested or determined be stated, and that the friction-time curves be referred to standard curves. Observations on the type, occurrence and duration of seizure and of the maximum friction are given. Friction-time curves for four kinds of oil are presented, including straight mineral oil types, acid types, sulphur type and chlorine type. The results obtained from a series of tests on the four-ball machine can be given in the form of a table, of which a typical example is presented.

The Oxidation of Transformer Oil. P. M. HEERTJES, C. D. TEN HAVE AND J. J. LOCK, *Jour. Inst. Petroleum* **33** (1947) pp. 552-73.

Bruckman has constructed an apparatus in which transformer oil can be aged in 100 hours to the same extent as in 15 to 20 years of normal use. This apparatus automatically records the dielectric losses occurring through and during the oxidation. Accelerated ageing is achieved by passing through the oil, at 100° C. and in the presence of copper as catalyst, a stream of dry carbon dioxide-free oxygen at the rate of 1650 ml per hour per litre of oil. During the process the oil is subjected to 10,000 volts A.C. and the dielectric losses are measured at the increase in the angle of loss

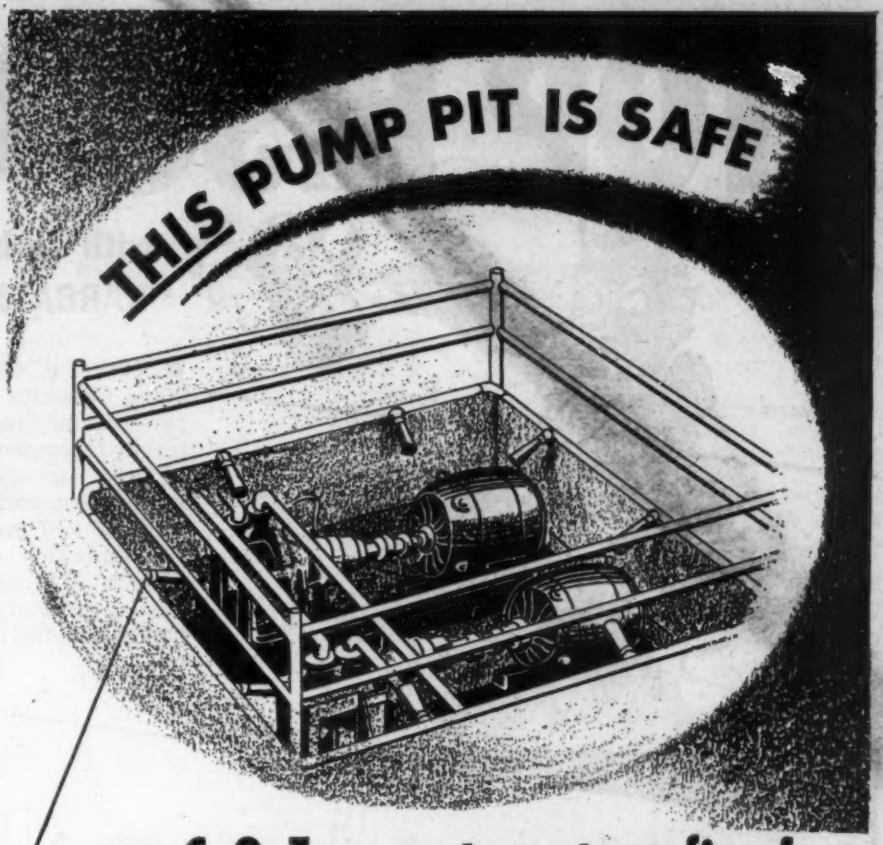
$tg\delta$, which is recorded automatically. The present research work was extended to include other gases, such as nitrogen, and to determine as far as possible the chemical changes in the oil. Acid saponification and hydroxyl numbers of the oil were measured, as well as the amounts of carbon dioxide and water, and volatile products formed in the course of the oxidation. The apparatus used is described in considerable detail, as well as the analytical procedure used. Data are presented in detail in tabular and graphical form. In general, it has been found that the compounds formed continue to react with each other after their formation, thus forming sludge probably combined with copper. All the data indicate that the reactions and the products are related to each other. The authors believe that hydroperoxides are formed by the oxidation of paraffinic side chains of the mainly napthenic-paraffinic molecules. The hydroperoxides oxidize easily, thus giving rise to the aldehydes or ketones, carbon dioxide and water. Some of the carbonyl compounds oxidize to acids, after which the hydroxyl compounds form esters with the acids.

Petroleum Products in the Printing Ink and Allied Industries. A. C. HEALEY, *Ph.D., M.Sc., F.R.I.C.*, AND L. IVANOVSZKY *(fellow), Jour. Inst. Petroleum* **33** (1947) pp. 531-51.

Printing inks consist of a colored pigment or dye dispersed or dissolved in a suitable medium. The purpose of the medium is to carry the pigment during the printing process from the ink duct or reservoir, via the printing plate, to the paper, and secondly, to fix the pigment to the paper after printing. Petroleum products are of considerable importance in printing ink manufacture, both as coloring materials in the form of carbon black and lamp-black, and as components of the vehicle. The methods by which fluid ink is converted into a rub-proof film are at first described and briefly discussed. The types of printing machines are considered. Among the petroleum products discussed in the article are volatile petroleum products, mineral oils, carbon black, petroleum waxes, petroleum jelly, bitumens, petroleum aromatic extracts and resins, naphthenates, greases, wetting agents, and miscellaneous products. The author concludes that there is no such thing in the printing industry as a standard formula. Different manufacturers use widely divergent formuli, using different materials to produce inks of approximately similar characteristics. As a result it is difficult in many cases to specify precisely qualities that are required in petroleum products.

Simplified Still Head with Automatic Control of Reflux Ratio. RICHARD KIESELBACH, *Ind. and Eng. Chem., Anal. Ed.* **19** (1947) p. 815.

The still head described is of the intermittent take-off type, in which the entire condensate is directed alternately to the receiver or to the column as reflux. The valve controlling this flow is operated by a commercial electric cycle timer. No lubricant is required on the valves exposed to the condensate and no stopcock is needed when the head is used at atmospheric pressure. When used under vacuum, one stopcock, exposed only to cold vapor, need be manipulated to withdraw condensate from the receiver.



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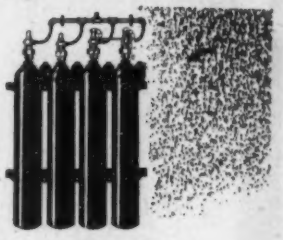

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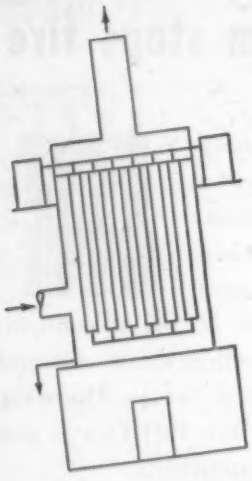

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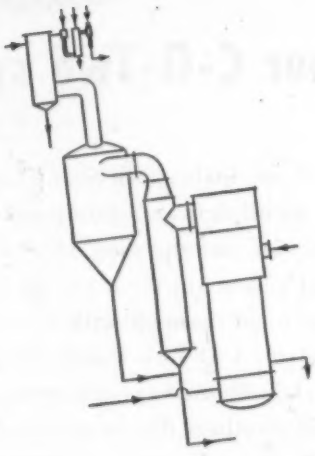

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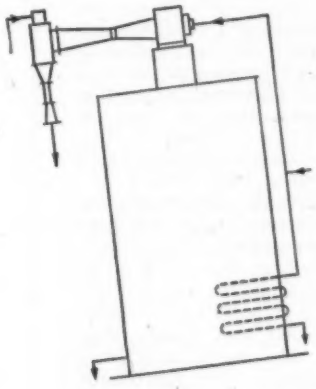

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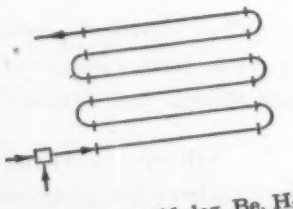

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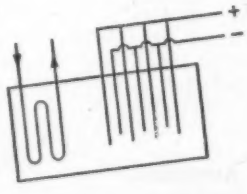

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Compiled by HEINZ HEINEMANN

REFINING

U.S.P. 2,427,589. Method of Refining Hydrocarbon Oil with a Sludge-Forming Reagent. V. L. Chechot to The Atlantic Refining Company.

A hydrocarbon oil which is to be refined is mixed with a sludge-forming reagent, such as H_2SO_4. The mixture obtained is supplied in the form of a film to an inclined or rotating surface heated to a temperature sufficient to coke the sludge resulting from the reaction of the reagent with the oil but insufficient to deleteriously affect the oil. The sludge is deposited and coked upon the heated surface. The oil is separately removed from this surface, e.g., by draining. The inclined or rotating surface can, for example, be heated to a temperature of 250-600° F., if H_2SO_4 is employed as the sludge-forming reagent.

U.S.P. 2,427,925. Diolefin Production. J. A. Sameniego and M. Souders to Shell Development Company.

A hydrocarbon mixture containing an acyclic diolefin, such as isoprene or piperylene or both, in admixture with hydrocarbons of a greater degree of saturation is contacted with SO_2 under conditions resulting in the formation of reaction products, predominating in the sulfone of the acyclic diolefin in admixture with SO_2 and unreacted hydrocarbons. The reaction products are passed from the reaction zone into a vaporizing zone where a vapor fraction comprising SO_2 and unreacted hydrocarbons is separated from a liquid fraction comprising the sulfone, SO_2 and residual unreacted hydrocarbons. The liquid fraction is stripped with SO_2 in a stripping zone to remove substantially all residual unreacted hydrocarbons. The diolefin sulfone freed of unreacted hydrocarbons is passed to a decomposition zone where it is heated at a temperature sufficiently high to decompose it with the liberation of the diolefin. The effluent from the decomposition zone is contacted in an extraction zone with a countercurrent stream of selective solvent for the diolefin. The diolefin is separated from resulting rich solvent containing the diolefin. SO_2 is passed from the extraction zone to the stripping zone.

U.S.P. 2,428,067. Treatment of Hydrocarbon Oil. G. H. Evans, F. T. Klopsch, and D. M. McCormick to The Atlantic Refining Company.

Viscous hydrocarbon oil is selectively extracted by nitrobenzene under addition of a small amount of ammonia and of an alkaline earth metal soap of a fatty acid sufficient to promote phase separation. Emulsion difficulties are avoided and good separation of the raffinate phase from the extract phase is obtained by the additives, and the possible throughput of the system is greatly increased. An amount equivalent to 0.005-0.02 gallon of 19° Bé. aqueous ammonia per barrel of oil and 0.0015-0.015 percent of the soap can, for example, be employed in the process.

U.S.P. 2,428,102. Separation of Beta-Ethylnaphthalene. W. Swietoslawski to Koppers Company, Inc.

A mixture of ethylnaphthalene isomers, as, for example, obtained by the catalytic alkylation of naphthalene with ethylene, is dissolved in a solvent the m.p. of which is lower than that of the eutectic mixture of the ethylnaphthalene isomers and which is non viscous at the crystallization temperature of beta-ethylnaphthalene. The solvent mixture is cooled to —40° to —50° C. Crystallized beta-ethylnaphthalene is separated from the mixture. An aliphatic alcohol, such as methanol or ethanol, is preferably employed as the solvent in an amount at least equal to the weight of the mixture of isomers to be treated.

U.S.P. 2,428,120. Recovery of Furfural from Furfural-Polymer Mixtures. G. H. Miller to The Texas Company.

A furfural feed containing a substantial amount of ploymer formed during the extractive distillation of olefin and di-olefin containing hydrocarbon mixtures with furfural at elevated temperatures is contacted with a relatively large proportion of water in the presence of a small amount of water soluble wetting agent effective to coagulate the polymer in free settling form, particularly an organic alkali metal sulfate or sulfonate. An aqueous phase comprising furfural dissolved in water and an insoluble phase comprising polymer are formed and separated.

U.S.P. 2,428,467. Two Stage Azeotropic Distillation of Toluene with Methanol. T. A. Petry, R. Lee, J. L. Biles, and W. V. Betts to Socony-Vacuum Oil Company, Inc.

A toluene containing petroleum fraction as, for example, obtained in vapor phase cracking of hydrocarbon oils, is first distilled to segregate therefrom a fraction of about 245° F. end point measured according to the specifications of the ASTM. This fraction which contains substantially all of the toluene originally present but the toluene content of which is less than about 40 percent by volume, is distilled in the presence of an added amount of methanol which is not less than about $M = .86(100-T) + .28T$ where M is the added methanol expressed as volume percent of the charge fraction and T is the volume percent of toluene in the charge fraction. The charge fraction is first distilled in the presence of a portion of the methanol to yield a first distillation residue containing substantially all of the toluene and having a toluene concentration of more than 40 percent. This residue is then distilled in the presence of the remainder of the methanol to yield as a second residue all of the toluene in a condition of relatively high purity of 90 percent or more.

U.S.P. 2,428,516. Process of Separating Isobutylene from Admixture of Butent-1 and Isobutylene. H. E. Drennan to Phillips Petroleum Company.

A liquid mixture of butene-1 and isobutylene, as obtained, for example, in the dehydrogenation of a mixture of butanes, is passed over an isomerization catalyst consisting of calcined brucite in a reaction zone maintained at a temperature of 200-400° F. and under a pressure sufficient to maintain liquid phase throughout the conversion zone. At least 90 percent conversion of butene-1 to butene-2 is effected with only little polymerization occurring. The liquid conversion products are removed from the reaction zone. The polymer formed, which is dissolved, is washed off the catalyst and carried away by the liquid feed and conversion products as fast as it is formed. The conversion products are fractionally distilled to separate the butene-2 from the isobutylene.

U.S.P. 2,428,521. Recovery of Ethane and Gasoline from Natural Gas. J. W. Latchum to Phillips Petroleum Company.

A natural hydrocarbon gas is contacted with an absorption oil at atmospheric pressure to produce a treated gas substantially free of high boiling natural gasoline boiling range hydrocarbons and containing some moisture, and a rich absorption oil containing the last named hydrocarbons. The extracted gas stream is treated with a dehydrating means. The resulting dry gas is chilled and contacted with a chilled absorption oil to produce gas substantially free from ethane and higher boiling hydrocarbons, and a cold rich absorption oil containing these materials. The treated gas is removed as one of the products of the process. The two rich absorption oils are combined and passed to a venting zone where dissolved gas substantially lighter than ethane is vented. The vented absorption oil is passed to a fractionating zone where an ethane concentrate and a natural gasoline product are separated as further products of the process. A portion of the remaining absorption oil is recycled to the first absorption stage, while the remaining portion of this oil is chilled and recycled to the second absorption stage. A straight run paraffinic

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oil boiling from about 200-350° F. can, for example, be employed as the absorption oil.

U.S.P. 2,428,611. Recovery of Azeotrope Former in Azeotropic Distillation of Hydrocarbons. C. H. O. Berg to Union Oil of California.

A complex hydrocarbon fraction is distilled in the presence of a sufficient amount of an azeotrope former to produce an azeotrope distillate consisting of at least one of the hydrocarbons of the complex fraction together with the azeotrope former. At least one of the remaining hydrocarbons of the complex fraction is left in the distillation residue. The azeotropic distillate is extracted with an agent adapted to dissolve one of the components only of this distillate. The solution formed is separated from the mixture of the components and undissolved portions of the dissolvable component. This mixture is distilled in a separate distilling zone to distill overhead substantially all of the dissolvable component and a portion of the other components, while the remaining portion of the other components is left in the residue. The overhead distillate is extracted in a separate extraction zone with a further amount of the same agent to dissolve a further quantity of the dissolvable component. The complex hydrocarbon fraction can thus be separated into component parts of dissimilar characteristics.

U.S.P. 2,428,668. Sulfuric Acid Olefin Absorption. H. J. Hibshman and I. P. Jones to Standard Oil Development Company.

An isobutylene-containing hydrocarbon gas is heated to a temperature of 80-200° F., and then contacted with H_2SO_4 of 85-98 percent concentration, heated before to the same temperature range, in an atomized state for a period of time of 0.2-3 seconds. Unreacted hydrocarbon gas is then immediately separated from the sulfuric acid. The sulfuric acid extract obtained is diluted with water to recover isobutylene, which can thus be obtained in a purity of 100 percent.

U.S.P. 2,428,686. Method of Treating a Hydrocarbon Feed Stock with a Solution Comprising Aryl Sulfonic Acid. G. H. Shipley, Jr., and G. W. Wilson, Jr., to Standard Oil Development Company.

A hydrocarbon feed stock is contacted with a solution comprising a major portion of aryl sulfonic acid and a minor portion of As_2O_5 dissolved in an organic solvent, such as an aromatic hydrocarbon, at a temperature of 100-400° F. The solution can contain, for example, 30-50 percent by volume of toluene sulfonic acid and 0.0013-0.013 gram mols of As_2O_5 per 100 grams of solution. The process is employed, for example, in the removal of undesirable constituents from hydrocarbons, e.g., in removing olefinic materials from toluene to prepare nitration grade toluene, in the desulfurization of hydrocarbons, and the like.

U.S.P. 2,428,690. Method of Treating Hydrocarbons. C. W. Tyson, D. L. Campbell, H. Z. Martin, and E. V. Murphree to Standard Oil Development Company.

A stream of low-boiling hydrocarbon distillates is passed in vapor form upwardly through a refining zone. A vertical column of solid refining agent in finely divided form is maintained in ad-

mixture with a fluidizing gas maintaining the refining agent in a freely flowing fluidized state throughout the length of the column. The finely divided refining agent is discharged from the base of the column into the hydrocarbon stream. The velocity of the vapors in the refining zone is controlled to maintain a relatively dense, turbulent mixture of vapors and refining agent. After purification of the vapors the refining agent is separated from these vapors and conducted to the top of a second vertical column of the finely divided refining agent where it is again maintained in admixture with a fluidizing gas. The refining agent is discharged from the base of this second column into a stream of gases separate and independent from the hydrocarbon vapors. This stream of gases with the refining agent contained therein is passed through a regenerating zone. After regeneration the refining agent is separated from these gases and conducted to the top of the first-named column. Naphtha, gasoline, kerosine, gas oil, or the like, can be refined by this process by means of clay, bauxite and other purifying materials.

U.S.P. 2,428,695. Hydrocarbon Extraction Process. D. C. Bond and M. Savoy to The Pure Oil Company.

A mixture of open-chain and closed-ring hydrocarbons is separated into a plurality of fractions at least one of which contains a higher proportion of open-chain hydrocarbons and at least one of which contains a higher proportion of closed-ring hydrocarbons than the original mixture. For this purpose, the mixture is contacted with at least 2 volumes of an aqueous solution containing not less than 10 percent by weight of free alkali metal hydroxide and not less than 10 percent of an alkali soap of naphthenic acids or petroleum partial oxidation products. The resulting mixture is separated into two layers. The hydrocarbon fractions rich in closed-ring compounds are recovered from the extract layer. A solventizer, such as cresol, can be added to the treating solution to keep the soap in solution.

CATALYST PREPARATION, ACTIVATION, AND REGENERATION

U.S.P. 2,428,741. Catalytic Conversion of Hydrocarbons. C. J. Plank to Socony-Vacuum Oil Company, Inc.

Hydrocarbon vapors are contacted at conversion conditions with a catalytic composition prepared by depositing silica on active alumina and thereafter treating the composite so produced with a mixture of HF and H_3PO_4 or an alcoholic solution of these acids. The new catalyst is particularly useful in the cracking of hydrocarbons causing very high yields of gasoline.

U.S.P. 2,428,524. Resolution of Aqueous Hydrofluoric Acid. M. P. Matuszak to Phillips Petroleum Company.

Hydrofluoric acid used as catalyst in the conversion, e.g., the alkylation, of hydrocarbons is purified by distillation under conditions such that an azeotropic mixture of water and HF is formed. This azeotropic mixture, which contains at least about 60 percent by weight of water, is distilled in the presence of at least one alkali metal fluoride dissolved in the kettle liquid. An overhead product comprising water and a liquid bottom product comprising the alkali metal fluoride and HF are removed from the distillation. The bottom product is stripped under conditions such that more concentrated HF is recovered as an overhead product of the stripping. The liquid bottom product from this stripping comprises the alkali metal fluoride and is passed to the distillation.

CRACKING AND REFORMING

U.S.P. 2,427,341. Catalytic Conversion of Hydrocarbons. J. G. Alther to Universal Oil Products Company.

Hydrocarbons are introduced to the lower portion of a vertically elongated reaction zone while hot regeneration catalyst is simultaneously supplied to the same portion of this zone from a regenerating zone. The resultant mixture is passed upwardly through the reaction zone under conversion conditions. The mixed conversion products and contaminated catalyst particles are discharged from the upper portion of the reaction zone into a separating zone disposed above the regenerating zone. The upward velocity of the mixture in the separating zone is reduced to separate contaminated catalyst from the conversion products. The separated catalyst is accumulated in a substantial body in the separating zone and is stripped there of occluded volatile hydrocarbons by passing an inert gas upwardly through this body at a velocity such as to maintain this body in a turbulent fluidized condition and to

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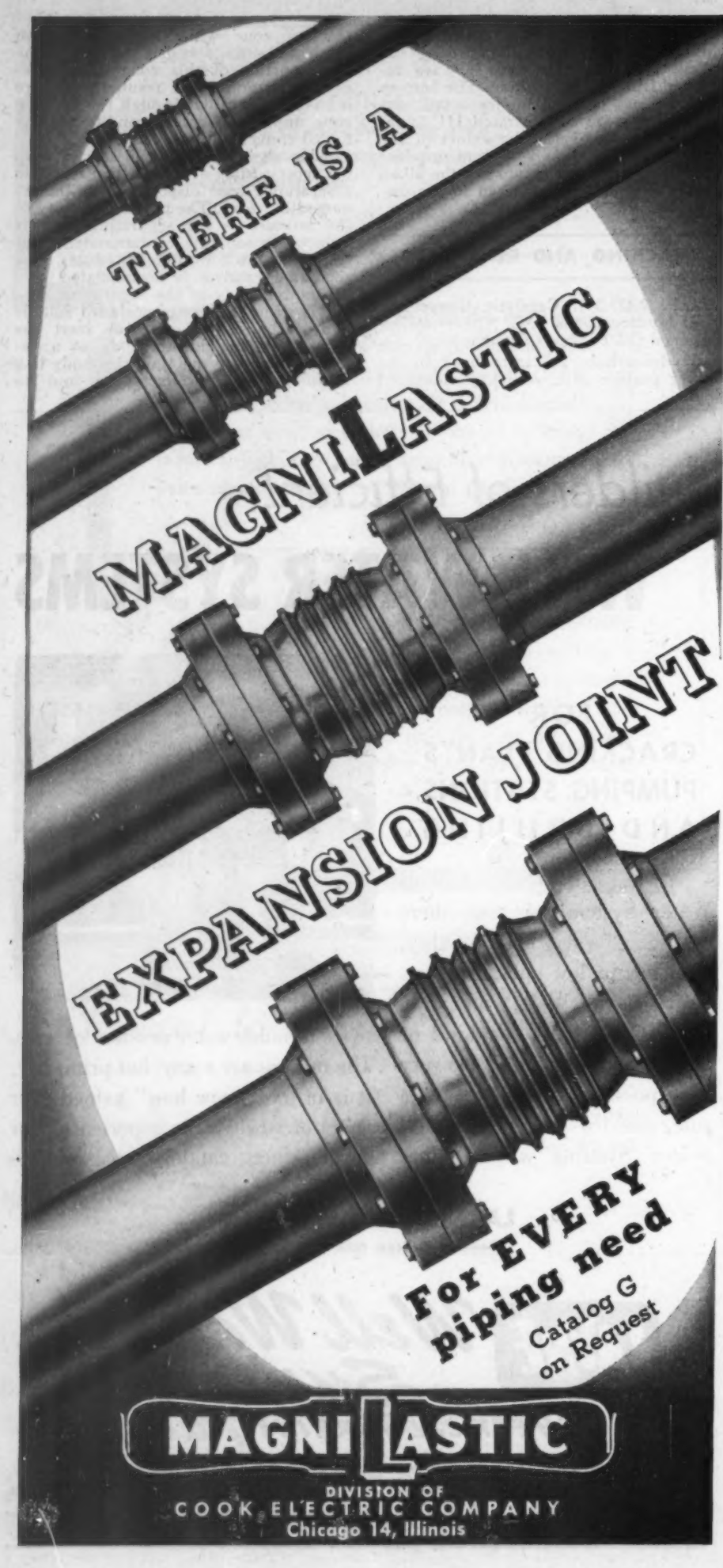

transport the catalyst particles in an upward direction within this zone. Contaminated catalyst is then passed downwardly from a point substantially above the bottom of the fluidized body in the separating zone to the regenerating zone. A turbulent fluid-like bed of catalyst particles is maintained in the regenerating zone while the contaminants are burnt from the catalyst. Hot regenerated catalyst is recycled from this bed to the lower portion of the reaction zone. The process is particularly designed for the catalytic cracking of hydrocarbon oil to produce an aviation base and/or motor gasoline of high-octane rating, a siliceous cracking catalyst comprising subdivided silica particles and composited metal oxides, such as alumina, chromia or magnesia being preferably employed in this case.

U.S.P. 2,427,800. Catalytic Reforming of Mixed Gasolines. W. J. Mattox to Universal Oil Products Company.

A mixture of a straight-run gasoline fraction and a thermally cracked gasoline fraction, wherein the former fraction is present to at least 50 percent of the total mixture and the latter fraction consists primarily of hydrocarbons boiling above 120° C., is subjected to the action of a reforming catalyst in the presence of hydrogen in a ratio of 0.5-15 mols of hydrogen per mol of hydrocarbon. A temperature of 400-600° C. and a pressure in excess of 30 atm. are maintained in the reaction zone. The cooled reaction products are introduced into a first separating zone where a gaseous mixture rich in hydrogen is separated from liquid hydrocarbons. The gaseous mixture is recycled to the reaction zone. The liquid hydrocarbons are introduced into a second separating zone maintained at a substantially lower pressure than the first separating zone. Further separation of gases containing hydrogen, is therein effected. A portion of the liquid hydrocarbons obtained from this second zone is recycled to the first separating zone. The remaining portion of the liquid hydrocarbons is heated and then introduced into a third separating zone wherein light gases are separated from the liquid hydrocarbons. The light gases from the second and third separating zones are introduced into an absorber wherein small amounts of gasoline hydrocarbons entrained in the gases are recovered. The liquid hydrocarbons withdrawn from the third separating zone are stabilized and form the reformed product desired which is appreciably improved in anti-knock properties.

U.S.P. 2,427,820. Catalytic Cracking Process. L. C. Thomas to Universal Oil Products Company.

A mass of finely divided solid cracking catalyst is maintained in a reaction zone. A gaseous heat carrying medium, such as a superheated light hydrocarbon gas, is introduced into the lower portion of this zone and passed upwardly therethrough in contact with the catalyst at sufficient velocity to fluidize the catalyst and form in this zone a lower region of relatively high density and an upper region of relatively low density. The heavy hydrocarbon oil to be cracked is simultaneously introduced, at least partially in liquid state, to the reaction zone above the point of introduction of the heat carrying medium. This medium is introduced at higher temperature than the oil in order to supply heat for the

cracking reaction. The heavy hydrocarbons are retained in contact with the catalyst for a sufficient time to effect substantial cracking. The resultant conversion products are removed from the reaction zone. The gaseous heat carrying medium can comprise hydrocarbon gases formed by the cracking of the heavy oil.

U.S.P. 2,428,257. Conversion of Hydrocarbon Oils. J. E. Ahlberg and C. L. Thomas to Universal Oil Products Company.

Hydrocarbon oil is contacted under conversion conditions with a catalyst composite comprising precipitated hydrated silica, precipitated hydrated alumina and precipitated hydrated zirconia substantially free of alkali metal compounds, and tripoli substantially free of alkali and alkaline earth metal compounds and of metal oxide impurities. Either the precipitated alumina or the precipitated zirconia can also be omitted. The process is particularly suited for the production of very high anti-knock gasoline by cracking petroleum oil, such as Pennsylvania or Mid-Continent gas oil. The catalyst is highly selective in prolonged use and is less disposed to accumulate carbonaceous deposits than other known catalysts.

U.S.P. 2,428,532. Catalytic Hydrocarbon Conversion Process in the Presence of Steam. W. A. Schulze and C. J. Helmers to Phillips Petroleum Company.

A stream of hydrocarbon oil boiling between 400-700° F. is introduced into a closed zone where it is cracked at a temperature of 900-1000° F. and in admixture with superheated steam to furnish endothermic heat of conversion, in the presence of a solid bauxite or alumina catalyst. The oil is passed through his zone at the rate of 2-8 barrels of oil per ton of catalyst per hour. Unconverted and relatively heavy hydrocarbons and a gasoline stock having a relatively high olefin content and boiling in the range of 150-400° F. are recovered from the cracking effluent. A stream of this gasoline stock is transmitted into a second closed zone where it is subjected to exothermic aromatization conditions at a higher temperature than in the cracking step, viz., in the range of 1050-1250° F., in admixture with steam diluent to control the exothermic aromatization in the presence of a solid bauxite or alumina catalyst. The gasoline stock is passed through this second zone at the rate of about 5 barrels per ton of catalyst per hour. A hydrocarbon material of enhanced anti-knock characteristics, of low olefin and high aromatic hydrocarbon content and boiling within the gasoline boiling range is obtained from the effluent of this treatment.

U.S.P. 2,428,666. Catalytic Conversion of Hydrocarbons. C. E. Hemminger to Standard Oil Development Company.

A relatively heavy hydrocarbon oil, e.g., a topped petroleum crude, is heated to vaporize at least a portion thereof. Powdered catalyst is added to the resulting vapors to form a suspension of catalyst in the vapors. This suspension is heated during a period of time short enough to prevent substantial cracking, for example for 3-5 seconds to a temperature of about 925° F. a second portion of powdered catalyst is then added to the suspension, whereupon the

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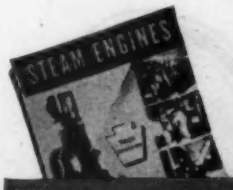

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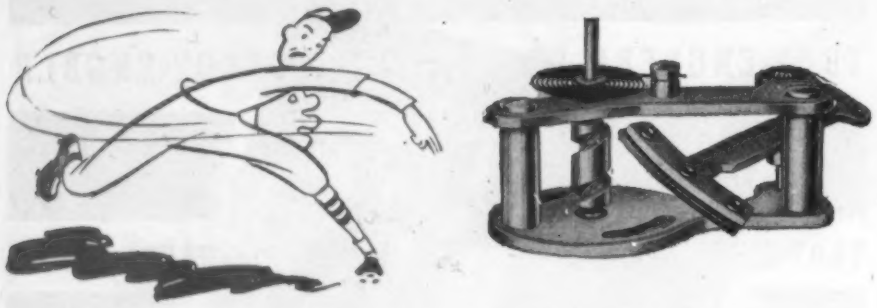

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resulting suspension is passed through a reaction zone where the vapors remain for a sufficient period of time to effect their conversion into hydrocarbon fractions boiling within the gasoline range, e.g., for 1 minute.

U.S.P. 2,428,691. Process for Stripping Spent Catalyst. C. W. Tyson to Standard Oil Development Company.

A stream of hydrocarbon oil vapor is passed upwardly through a conversion zone containing finely divided catalyst at a velocity controlled to maintain a relatively dense, turbulent layer of catalytic material in the bottom portion of this zone. Vaporous conversion products removed from the upper portion of the zone are fractionated to separate a desired product therefrom. A stream of catalytic material containing appreciable amounts of vaporous conversion products and solid combustible deposits is continuously withdrawn from the bottom portion of the conversion zone and passed downwardly through a vertical column maintained substantially full of this catalytic material. A stream of air is introduced into the bottom of this column. The amount of air is controlled so as to cause its rapid complete consumption to form a combustion gas substantially devoid of free oxygen. This combustion gas is passed upwardly through the column to effect replacement of vaporous conversion products from the catalyst. The catalyst is passed from the base of the column into a regeneration zone where it is subjected to further regeneration to remove additional amounts of combustible deposits, and is then returned to the conversion zone. The process can, for example, be used in cracking of gas oil or other hydrocarbon oils or in the reforming or refining of a light distillate stock such as naphtha.

U.S.P. 2,428,715. Catalytic Cracking of Hydrocarbons. M. M. Marisic to The Pure Oil Company.

Mineral oil is contacted with a siliceous catalyst under suitable conditions for cracking a substantial portion of the oil to gasoline, which is separated. The remaining insufficiently cracked stock is contacted with activated carbon under suitable cracking conditions. The activated carbon shall be capable of cracking at least 10 percent of East Texas gas oil with an API gravity of 37 and a boiling range of about 430-700° F. to gasoline at 850° F. when this oil is cracked in the presence of the activated carbon in a weight ratio of 1 p. of oil to 0.575 p. of catalyst over a 30-minute period. Spent carbon is separated from the effluent from the cracking zone, and this effluent is then separated into high octane gasoline and insufficiently cracked heavier hydrocarbons. The latter are recycled to the carbon cracking zone. The spent carbon catalyst is regenerated and recycled.

ISOMERIZATION

U.S.P. 2,427,775. Isomerization of Saturated Hydrocarbons. B. S. Friedman to Universal Oil Products Company.

An isomerizable saturated hydrocarbon, such as n-pentane, is contacted under isomerizing conditions with a catalyst comprising a Friedel-Crafts type metal halide and a hydrogen halide in the presence of 0.1-10 percent by volume, based on the hydrocarbon treated,

of a mono-chloro or mononitro substituted aromatic hydrocarbon, e.g., mononitrobenzene or monochlorotoluene. These aromatic compounds selectively suppress the decomposition of the isomerizable hydrocarbons and permit the isomerization of the saturated hydrocarbons with a high degree of efficiency and low catalyst consumption.

U.S.P. 2,428,455. Isomerization of Olefins. A. H. Friedman to Phillips Petroleum Company.

A 1-olefin with at least 4 C atoms in a chain is passed over a solid catalyst comprising TiF_4 at a temperature of 100-500° F. and at atmospheric pressure. The carbon to carbon double bond is shifted by this process to a more central position. For example, low-boiling isohexenes are obtained by this process from vapors of hexene-1.

ALKYLATION

U.S.P. 2,428,417. Catalytic Alkylation of Iso-Paraffin with Olefins. W. W. Gary to The M. W. Kellogg Company.

A high-octane motor fuel is prepared from a mixture of isobutane, n-butane and butylene in which there is less than an equivalent proportion of butylene relative to the isobutane. Butylene is added to this mixture which is then subjected to catalytic alkylation. N-butane is separated from the butylene-isobutane alkylation product. This n-butane is converted into isobutane by catalytic isomerization. The isobutane thus obtained is pyrolized to produce butylene which is added to further quantities of the feed mixture to be alkylated. In a corresponding manner n-butane alone can first be isomerized and a portion of the isobutane obtained then pyrolized to form butylene which is alkylated with another portion of the isobutane formed before.

U.S.P. 2,428,279. Alkylation of Aromatics. V. N. Ipatieff and C. B. Linn to Universal Oil Products Company.

An aromatic hydrocarbon, such as benzene, is reacted with an alkylating agent, such as ethylene, propylene or butylene, under alkylating conditions in the presence of BF_3 and fluorosulfonic acid. The alkylation can be effected at a temperature of 0-100° C. While the two compounds comprising the catalyst will no catalyze individually the interaction of say benzene and ethylene, a very satisfactory, liquid catalyst is formed by their combination.

U.S.P. 2,428,506. Alkylation Process. J. H. A. P. Langen Van der Valk to Shell Development Company

In the alkylation of an isoparaffin with an olefin or the like in the presence of conc. H_2SO_4 an emulsion of hydrocarbon and acid is circulated in a closed circuit under alkylation conditions. A part of this circulating emulsion is continuously withdrawn and contacted with isoparaffin substantially free from alkylating agents under conditions at which the acid content of the emulsified mixture is separated from at least a part of the reaction product while being maintained in intimate emulsified contact with the isoparaffin. The thus separated acid emulsion is returned to further alkylation.

U.S.P. 2,428,753. Process for Treating a Hydrocarbon Mixture which is Contaminated by Small Amounts of Organic Fluorine Compounds. C. B. Linn to Universal Oil Products Company.

Hydrocarbons as obtained by the alkylation of isoparaffins with olefins or with alkyl fluorides in the presence of active fluoride catalysts, such as HF or mixtures of HF and BF_3, often are contaminated by small amounts of organic fluorine compounds. Such hydrocarbon material is treated with fluosulfonic acid at defluorinating conditions of temperature and pressure, e.g., at a temperature of —50 to 150° C. and at a pressure sufficient to maintain the hydrocarbons in liquid phase. The purified hydrocarbon material is recovered.

U.S.P. 2,428,754. Process for Treating a Hydrocarbon or Hydrocarbon Mixture which is Contaminated by Small Amounts of Organic Fluorine Compounds. C. B. Linn to Universal Oil Products Company.

The process of this patent is similar to that of U. S. patent 2,428,753. Chlorosulfonic acid is employed here as the defluorinating agent.

DESULFURIZATION

U.S.P. 2,427,988. Process for Removing Sulphur Compounds from Hydrocarbons. P. H. Wilson, Jr., J. H. Wells, and P. M. Sommerfeld to Carnegie-Illinois Steel Corporation.

Thiophene is removed from aromatic hydrocarbons, such as benzene, by fractionating the hydrocarbons in the pres-

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ence of an alkali-water-alcohol solution containing NaOH or KOH in at least equimolecular amounts with the thiophene present, and enough of an aliphatic alcohol to carry over the aromatic hydrocarbons in the form of its azeotrope with the alcohol. The b.p. of the alcohol shall be below that of the hydrocarbon to be separated. The process can be employed generally for removing organic sulfur compounds present in coal tar and coke oven light oils from aromatic hydrocarbons.

HYDROGENATION, DEHYDROGENATION, AROMATIZATION

U.S.P. 2,427,791. Hydrogenation of Halogenated Hydrocarbons. V. N. Ipatieff and L. Schmerling to Universal Oil Products Company.

A halogenated saturated hydrocarbon, such as a mono- poly- haloalkane or a halocycloparaffin, is reacted with hydrogen in the presence of a hydrogenating catalyst, such as Ni, and an alcohol present in at least an equal mol. proportion with the halogenated hydrocarbon. The treating temperature shall not be less than 125° C. Under these conditions the alcohol simultaneously reacts with the hydrogen halide liberated by the hydrogenation reaction to form the corresponding organic halide which is stable under the reaction conditions and nonpoisonous to the hydrogenating catalyst. Saturated hydrocarbons are obtained.

U.S.P. 2,427,865. Conversion of Naphthenes to Paraffins. A. P. Lien and B. L. Evering to Standard Oil Company (Indiana).

A charging stock containing a substantial proportion of naphthenic and aromatic hydrocarbons is contacted with hydrogen and a body of liquid HF-BF_3 catalyst containing 1-40 percent by weight of BF_3, based on the HF, under conditions to hydrogenate naphthenic hydrocarbons. A liquid catalyst layer containing aromatic hydrocarbon is removed from the reaction zone. Dissolved and loosely-bound catalyst is removed from this layer, whereupon aromatic hydrocarbons are removed from this layer. A catalyst complex containing firmly-bound HF and BF_3 is subsequently separated from this layer. This complex is thermally decomposed to regenerate HF and BF_3. The catalyst separated from the catalyst layer is recycled to the reaction zone. A temperature of 150-450° F. and a total pressure of 600-3000 psi. can be maintained for the treatment of naphthenes in the reaction zone for a period of 0.01-4 hours. Paraffinic hydrocarbons are continuously obtained by the process.

U.S.P. 2,428,151. Process for Conversion of Hydrocarbons. F. E. Frey to Phillips Petroleum Company.

An aliphatic C_3-C_5 paraffin hydrocarbon, e.g., isobutane, is preheated to a temperature between 800-1200° F. at which catalytic dehydrogenation is practicable but at which noncatalytic conversion is practically negligible. The heated hydrocarbon is passed through a single massive stationary bed of dehydrogenation catalyst, such as granular bauxite. This catalyst bed is well-insulated to prevent gain or loss of heat through the walls of the converter, so that the only addition or withdrawal of heat occurs with the incoming and effluent streams. The contact time is such as to effect a conversion of 5-15 percent of the original hydrocarbon to less-saturated aliphatic hydrocarbons. The resulting mixture cooled because of the endothermic nature of the conversion is subjected to noncatalytic thermal dehydrogenation wherein sufficient heat is supplied to maintain the mixture at a temperature between 1100-1300° F. at which temperature noncatalytic thermal conversion to less-saturated hydrocarbons occurs at a practical rate. The mixture is held at this temperature for a time sufficient to effect additional dehydrogenation. The total conversion of the original hydrocarbon is increased to 15-40 percent.

U.S.P. 2,428,624. Method of Dehydrogenating Butene. C. H. Holder to Standard Oil Development Company.

A preheated mixture of steam and butene is charged to a reaction zone which contains a catalyst consisting essentially of MgO, Fe_2O_3, a stabilizer and a volatile alkali metal compound, such as K_2CO_3. The reactant is maintained at dehydrogenation temperatures and the activity of the catalyst is prolonged by the addition of an amount of CO_2 representing a small fraction of the total charge. The activity of other dehydrogenation catalysts can also be prolonged by the addition of CO_2 and hydrocarbons other than butene can also be treated by this method. CuO can, for example, be used as the catalyst stabilizer.

U.S.P. 2,428,692. Production of Isobutane and Isopentane by Destructive Hydrogenation of Petroleum Oils. A. Voorhies, Jr., to Standard Oil Development Company.

Petroleum hydrocarbons boiling at 100-800° F. are destructively hydrogenated in the presence of hydrogen and of a catalyst containing Ni, Co, W, or Mo and having hydrogenating and isomerizing properties. Isobutane and isopentane formed in this treatment are removed. Hydrocarbons with a boiling range of about 160-225° F. are recycled. The treatment can, for example, be effected at a temperature of about 600-800° F. and a pressure of about 1000-5000 psi.

POLYMERIZATION AND CONVERSION

U.S.P. 2,427,907. Polymerization of Olefin Hydrocarbons. H. Heinemann and W. A. LaLande, Jr., to Porocel Corporation.

Isobutylene is polymerized to liquid hydrocarbons by contacting it at a temperature of 350-600° F. with bauxite which has been activated by heating at a temperature of 700-1800° F. to a residual moisture content of not more than 6 percent by weight. The polymerization reaction can be carried out selectively in the presence of n-butylene.

U.S.P. 2,427,954. Copolymerization of Butenes and Propene. F. E. Frey to Phillips Petroleum Company.

A hydrocarbon mixture comprising propene and butenes in a molecular ratio

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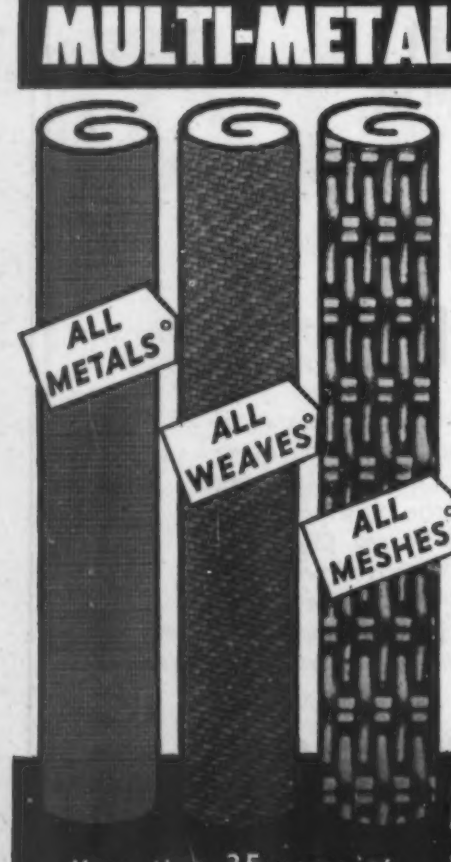

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of less than 1:1, as obtained for example by thermal dehydrogenation of isobutane, is divided into at least two portions. A recycle hydrocarbon mixture comprising propene is added to one of these portions to provide a mixture having a molecular excess of propene over butenes. This mixture is subjected to the action of an olefin polymerization catalyst at a temperature of 75-600° F. to copolymerize and convert butenes and propene into liquid hydrocarbons. Other portions of the first named hydrocarbon mixture are added at a plurality of points to the mixture undergoing polymerization. A hydrocarbon fraction comprising the normally liquid hydrocarbons produced is separated from the effluents of the polymerization zone. Another fraction comprising unreacted propene is also separated from these effluents and returned as the recycle hydrocarbon mixture of the polymerization reaction.

HEAVY OILS AND WAXES

U.S.P. 2,426,540. Process for Producing Hydrocarbon Oil Concentrates of Barium Sulfonates. F. M. Watkins and T. A. Hack to Sinclair Refining Company.

The hydrocarbon oil concentrates of barium sulfonates produced by this process are particularly valuable for blending with lubricating oils in which barium sulfonates act as corrosion inhibitors. For the production of these concentrates an intimate mixture of sulfuric acid-treated hydrocarbon oil (e.g. a Mid-Continent crude base distillate) and anhydrous Ba $(OH)_2$ having a particle size not coarser than 100 mesh is formed at a temperature below 135° F. The proportion of Ba $(OH)_2$ shall be in excess of that required to completely neutralize the acid oil. The mixture is agitated for a period not exceeding 30 minutes with 2.2-7 percent of water. The aqueous mixture is allowed to settle. Two layers are formed. The lower layer which has a relatively high concentration of barium sulfonates is separated from the upper layer which has a relatively low concentration of them.

U.S.P. 2,426,549. Lubricant Composition. W. J. Coppock to Sun Oil Company.

A mineral oil lubricant composition contains a small proportion of thymol effective to inhibit oxidation, and a small rust-preventive proportion of a substance comprising a substituted 1,2-dihydroxy-benzene having as the substituent a saturated aliphatic hydrocarbon group attached to the benzene ring in the para position and containing 4-12 C atoms.

U.S.P. 2,426,852. High Temperature Lubricant. D. L. Wright to Jasco, Inc.

The high temperature lubricant for lubrication of moving metal parts at temperatures above 500° F. comprises at least 60 percent by weight of mineral seal oil with a relatively low boiling range (e.g. about 510-670° F.), about 0.3-20 percent of polylauryl alpha methacrylate of an evarage mol. wt. of 5000-20,000, and about 0.05-2 percent of colloid graphite. The amount of the polymer used shall be sufficient to make a mineral oil solution with a viscosity of at least 100 seconds Sayboldt at 100° F. The primary purpose of using a low-boiling mineral oil is that it evaporates at high temperatures without formation of varnish or tar and without carbonization.

U.S.P. 2,427,272 Mineral Oil Composition. E. W. Fuller, H. E. Redman, and H. G. Berger to Socony-Vacuum Oil Company, Inc.

The mineral Oil composition comprises a major proportion of a viscous mineral oil fraction and an admixed small amount, sufficient to enhance the stability of this oil fraction, of an oil-soluble, acidic, phosphorus- and sulfur-containing reaction product obtained by reaction at a temperature of 50-125° C. of P_2S_5 and an oxidized paraffin wax characterized by a neutralization number of about 5-60 and hydroxyl and saponification numbers of at least 20. Instead of this acidic reaction product, an ammonium salt or a metal salt thereof can also be used. An oxidized paraffin slackwax meeting the indicated specifications can, for example, be used in the production of the additive. The formation of acidic oxidation products and sludge in the mineral oil is prevented by the additive and corrosion of metal engine parts is inhibted thereby.

U.S.P. 2,427,501. Oil Compound. M. P. Kleinholz to Sinclair Refining Company.

A petroleum lubricating fraction is compounded with minor proportions of alpha, alpha'-thidilauric acid and alpha-hydroxylauric acid. The relative proportions of these two additives shall be in the range of about 80-50 parts of the first named to about 20-50 parts of the second named substance. The total of the two

For HOT and HEAVY fluids pack with HUSKY J-M No. 5

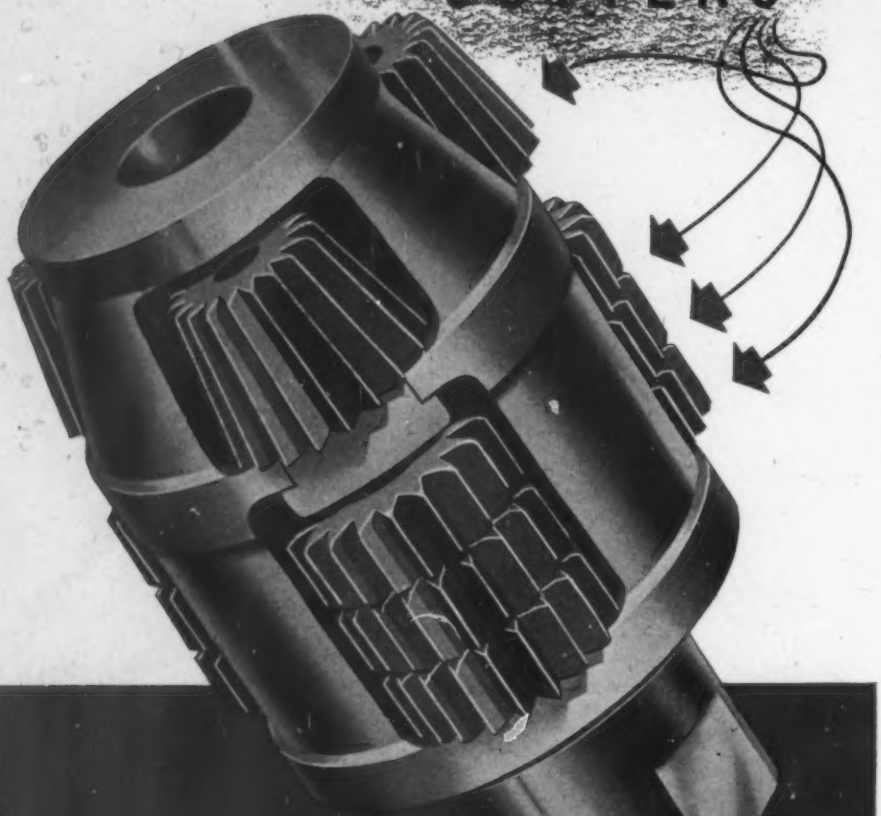
FOR LONG, HARD SERVICE against hot crudes and superheated steam—you can rely on J-M No. 5 Rod Packing. The plaits of copper-inserted asbestos yarn retain their lubricant in the toughest pumping . . . hold together on the start-ups when oil is congealed . . . resist glazing during running.

J-M No. 5 maintains its tight seal with minimum friction. Use it on rods, plungers and valve stems when the steam is over 500 F., the oil over 350 F. Get J-M No. 5 in coil or ring form from your Supply House—or write Johns-Manville, Box 290, New York 16, N. Y.

JOHNS-MANVILLE JM PRODUCTS

Johns-Manville PACKINGS & GASKETS

AIRETOOL'S NEW FORM* CUTTERS



...guarantees FASTER, MORE EFFICIENT TUBE CLEANING!

Shown above is the single expansion P-type cutter head with 3 cylinder cutters to the row. This same head is also furnished with two or four cylinder cutters, depending upon the coke condition. *This is only one of many types of Airetool cutter heads.* Single expansion, double expansion and wing-arm types are available to meet every refinery tube cleaning problem.

RECOMMENDED FOR STRAIGHT TUBES 2 1/4" to 4 3/4" I. D.

All of the cutters employed in Airetool cutter heads are of the *New Form* type which are proven to be *Absolutely Non-Tracking* . . . even after repeated cleaning operations.

AIRETOOL
MANUFACTURING COMPANY
316 S. Center St.
SPRINGFIELD, OHIO

WRITE FOR
BULLETINS ON ALL TYPES OF AIRETOOL REFINERY CLEANERS, EXPANDERS, AND ACCESSORIES.

additives is preferably in the range of about 0.01-0.2 percent of the lubricating oil. Rusting of metal parts is retarded by the additives.

U.S.P. 2,427,766. Compounded Lubricating Oil. H. Diamond to Shell Development Company.

The oxidation stability of a hydrocarbon lubricating oil composition which initially contains less than 6 percent aromatics is increased by the addition of an oil-soluble halogen free organic sulfide combined with 0.2-15 percent of a highly oil-soluble condensed ring polynuclear aromatic hydrocarbon, such as retene, alkyl naphthalene, or alkyl anthracene. The organic sulfide can, for example, be di- or poly-sulfide of a wax and can be employed in amounts of 0.001-1 percent.

U.S.P. 2,428,082. Oil Bleaching Process and Apparatus. R. R. King, S. E. Pack, and F W. Wharton to Mrs. Tucker's Foods, Inc.

Oil, e.g., heavy petroleum lubricating oil, is mixed in the liquid or completely liquefied state with an adsorbent, and the mixture is passed into the upper part of a deaerating zone maintained under vacuum and at a temperature substantially below effective bleaching temperature. The mixture is splattered there to finely divide it in the form of a spray so as to provide for rapid removal of air therefrom. Thereafter, the mixture is fed through a second zone under vacuum at an effective bleaching temperature, e.g. at 300-750° F. The oil is thus bleached with the adsorbent in the absence of air, so that oxidation of the oil and gum formation at bleaching temperature is prevented.

U.S.P. 2,428,123. Soda Base Grease. J. D. Morgan and R. E. Lowe to Cities Service Oil Company

Hot mineral lubricating oil containing at least 20 percent of refined Mid-Continent base stock is mixed with 10-30 percent by weight of sodium stearate. The resulting mixture is heated to a final temperature of 550-600° F. and then quickly chilled. A sodium base lubricating grease is obtained which is highly resistant to dissolution in water and contains 90-70 percent by weight of the mineral lubricating oil.

U.S.P. 2,428,340. Composition of Matter, Its Preparation and Utilization. P. M. Ruedrich to Griffin Chemical Company.

A mixture of lithium soap, e.g. lithium stearate, and an oxidized petroleum hydrocarbon is heated to form a solution. This solution is added to a mineral oil of lubricating viscosity heated to a temperature below its flash point. The lithium salt shall be present in the oil in an amount sufficient to substantially thicken the oil. The oxidized petroleum hydrocarbon shall have a flash point above that of the lubricating oil and shall be soluble therein. The use of oxidized petroleum hydrocarbon facilitates the incorporation of the lithium compound into the lubricant.

U.S.P. 2,428,623. Method of Regenerating Spent Caustic Alkali Solution. A. P. Hewlett and H. C. Paulson to Standard Oil Development Company.

Spent caustic alkali solution used for the treatment of naphtha is contacted with a petroleum hydrocarbon solvent oil boiling between 500-800° F. Mercaptides contained in the spent solution are extracted and the alkali solution is thus regenerated.

PETROCHEMICALS

U.S.P. 2,427,704. Process for the Production of Butadiene. J. E. Bludworth and M. O. Robeson to Celanese Corporation of America.

A mixture of tetramethylene oxide and water is preheated to at least a vaporizing temperature. The vapors are subjected at a temperature of 450-485° F. to the action of a dehydrating catalyst composed of mono-calcium and monosodium phosphate admixed with phosphoric acid and butylamine phosphate. Butadiene is obtained in high yield with a minimum formation of polymers.

U.S.P. 2,428,119. Preparation of Oxygenated Compounds from Olefins. C. G. Ludeman to Texaco Development Corporation.

Normally gaseous olefins are passed in contact with conc. H_2SO_4 having an organic free acidity of at least 90 percent in an absorption zone. An acid mixture containing dialkyl sulfate and alkyl sulfuric acid is thus formed wherein the dialkyl sulfate amounts to at least 30 percent by weight of the mixture. The acid mixture is contacted with an amount of anhydrous alcohol which is substantially the molar equivalent of the dialkyl sulfate content of the mixture and under conditions such that the dialkyl sulfate and the alcohol react to form an aliphatic ether which is separated from the alkyl sulfuric acid. A portion of this residual alkyl sulfuric acid is returned to the absorption zone. Another portion of the residual acid is treated in a hydrolizing zone with water to convert it into alcohol and produce spent acid. The alcohol is removed. The proportion of residual alkyl sulfuric acid mixture returned to the absorption zone is adjusted so that there is substantially no net production of alcohol over that required for reaction with the dialkyl sulfate.

U.S.P. 2,428,590. Production of Allyl Type Compounds. E. C. Shokal and T. W. Evans to Shell Development Company.

Propylene is cracked with selenium dioxide dissolved in an aliphatic carboxylic acid, such as acetic acid, at a temperature near the b.p. of the mixture, e.g., at about 95-130° C. The corresponding allyl ester, e.g., allyl acetate, is obtained. By hydrolization of the resulting, neutralized ester, allyl alcohol is obtained.

U.S.P. 2,428,727. Carbon Disulfide Process. C. M. Thacker to The Pure Oil Company.

Hydrocarbons are contacted with sulful in a conversion zone under conditions such as to yield CS_2 and H_2S as the chief reaction products. The reaction products are cooled sufficiently to condense unreacted molten S which is separated and recycled. The remaining reaction products are contacted with water and SO_2 at a temperature below 100° C., the amount of SO_2 employed being about the stoichiometric equivalent of the H_2S, and the water being less than the amount required to give a saturated solution of the SO_2. The suspended sulfur formed is separated from the aqueous medium and recycled to the conversion zone. CS_2 is separated from the aqueous medium and from the remaining reaction products. Excess SO_2 from the reaction gases is recovered in a partion of the aqueous medium freed from sulfur, and this aqueous medium is then recycled to the S-forming step.

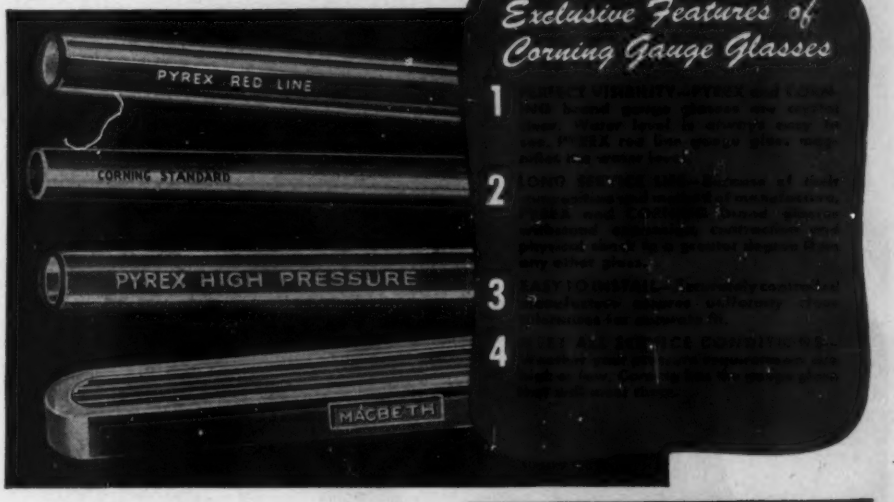
U.S.P. 2,428,738. Polyalkyl Aromatic Insecticides. R. R. Dreisbach and F. W. Fletcher to The Dow Chemical Company.

The insecticidal composition according to this patent comprises a solution in a petroleum distillate of a high-boiling fraction of the polyalkyl aromatic reaction product obtained by heating naphthalene and diethyl-benzene with a Friedel-Crafts catalyst. The fraction used shall boil above 230° C. at 25 mm. pressure, have a sp.g. of 1.00-1.05 and a refractive index of 1.60-1.63. Instead of the petroleum distillate another inert carrier for the insecticide can be used.

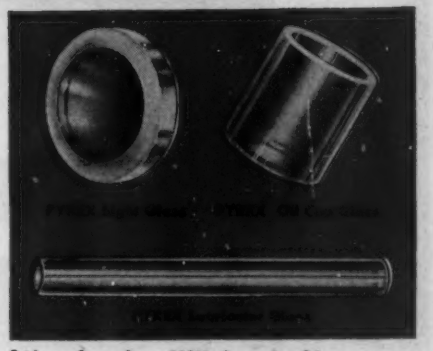
U.S.P. 2,428,745. Alkylation of Phenols. G. H. Stillson to Gulf Research & Development Company.

Monohydric phenols, such as cresols, are treated in the presence of a catalytic amount of tetraphosphoric acid with an olefin containing at least 3 C atoms, such as isobutylene, at a temperature of about 30-70° C. Alkylated phenols, such as tertiary butyl meta-cresol or 2,6 ditertiary-butyl-4-methyl phenol are obtained. They are useful as antioxidants, for example, in the prevention of oxidational deterioration of petroleum hydrocarbons.

BE SURE YOU HAVE EXTRA GAUGE GLASSES
...be sure they are made by Corning!



When you need gauge glasses, you need them quick. Low water in your boiler, for example, could cause damage costing real money. But a broken gauge glass wouldn't reveal the danger. There isn't any economy in being without spares. There just isn't any economy in being without Corning products. They provide the most dependable service at the lowest possible cost.



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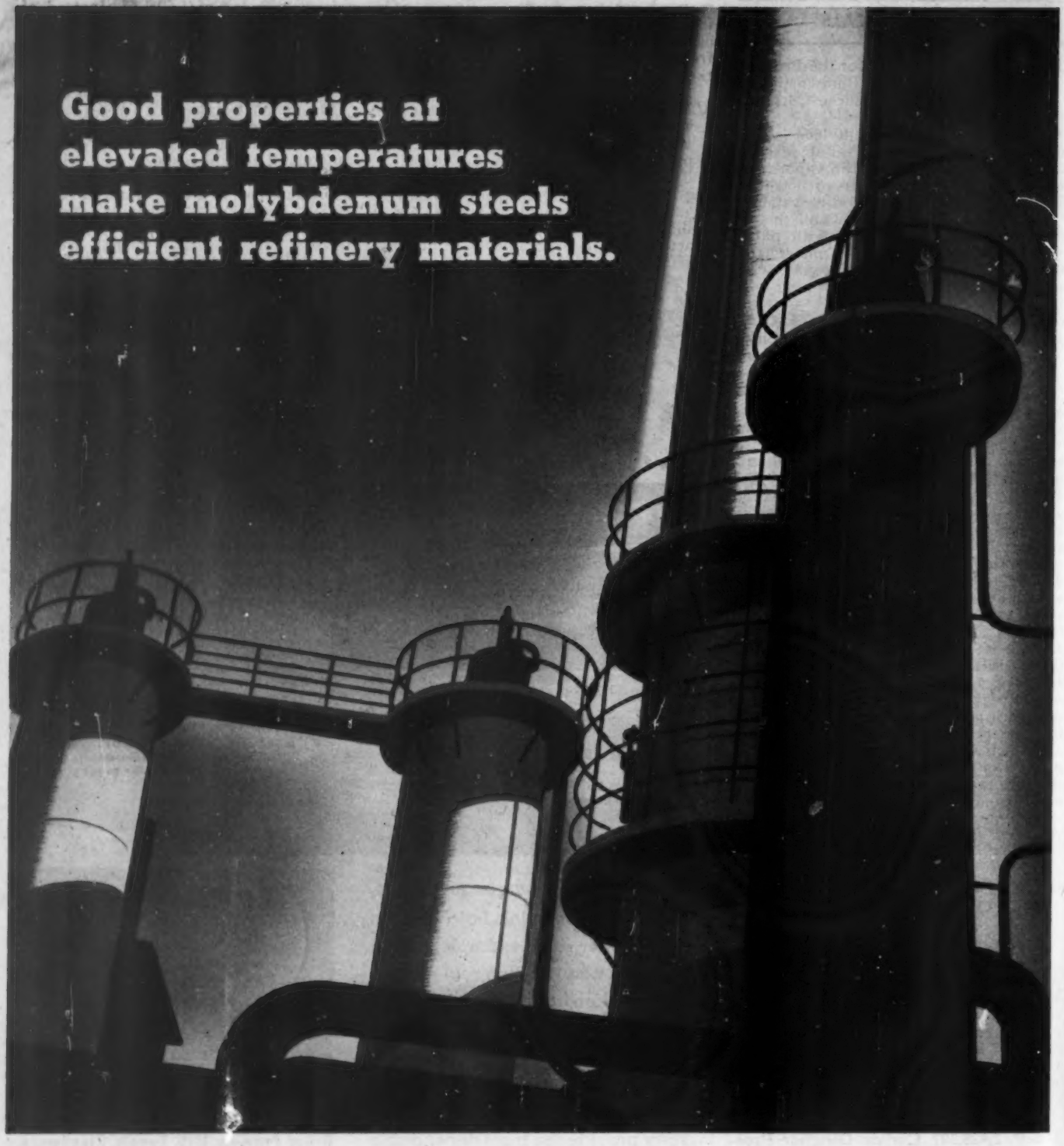

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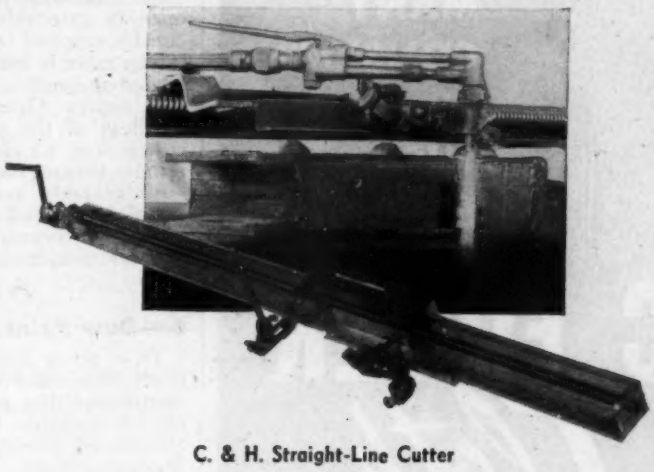
MOLY

NEW EQUIPMENT
. . . MANUFACTURERS' · LITERATURE . . .

For Copies of Manufacturers' Literature or More Information on Products Described on These Pages, Use the Number Shown for Each Item and the Address Listed

1—Welding Torch Holder

J. A. Campbell Company, 645 East Wardlow Road, Long Beach 7, Calif. is manufacturing a portable straight cutter which holds a welding torch and feeds it along a straight line, thus eleminating a wobbly, ragged edge. By turning a geared crank at the end, the torch travels along a pre-determined line and cuts through the metal as true as if lathe-cut.



C. & H. Straight-Line Cutter

Known as the C. & H. straight-line cutter the device will hold any torch. It has an adjustable holder so the torch can be angled to cut bevel or scarf. It has a positive feed so that the operator can control the torch speed. The torch is kept at a uniform distance from the work. It makes perfectly square cutoffs of plates, rods and bars.

This portable cutter is five feet in length and weighs only 42 pounds and is so ruggedly built it will not distort or become damaged from rough handling.

2—Hand-Held Electric Tachometer

A new hand-held electric tachometer which weighs only three pounds has been announced by General Electric Company, Schenectady 5, N. Y. It is designed to give accurate and direct readings of linear speeds from 10 to 10,000 rpm, and of rotational speeds from 100 to 10,000 rpm. Using accessories, rotational speeds from 10 to 100,000 rpm may be measured.

The tachometer comprises two units —the head, which is placed in contact with the moving object, and the indicating unit attached by a flexible electric cable. Speed ranges can be changed while the spindle is rotating. Accurate speed indications are assured by a low driving torque of only ¼ ounce-inch. The instrument cannot be damaged by overspeeding. Ask for Bulletin GEA 4895.

Hand-Held Electric Tachometer

3—Special Valves

A series of non-lubricated selector and shut-off valves especially adapted to chemical and process service is announced by The Parker Appliance Company, 17325 Euclid Avenue, Cleveland, Ohio.

Handling air, gases or hot liquids up to 400° F., and at non-shock pressures as high as 500 psi., the valves are an adaptation of the Parker aircraft types. The valves, available in a variety of metals, and designed with non-lubricated rotors of inert graphite, are especially suited to the handling of corrosive fluids.

The shut-off valves are constructed around a cylindrical graphite rotor which has a central passage at right angles to the cylindrical axis. When the passage is lined up with the ports of the valve body, straight-through flow is permitted. Indexed 90°, blank surfaces of the cylinder wall face the ports. The rotor is in two parts, and contains four springs which exert a separating pressure at right angles to the passageway, to provide initial contact of the rotor parts against the valve body.

The selector valves come in a variety of rotor arrangements, permitting the inter-connection of various combinations of as many as five separate lines. An "all-off" position also is provided.

Parker Appliance Company's New Shut-off Valve.

Positive identification of new equipment bulletins by name and number listed in Petroleum Refiner insure prompt replies to your requests for literature and information.

4—Welding Equipment

To meet the demand for equipment suitable for the high speed, high quality welding of the light metals such as stainless steel, aluminum and magnesium the Air Reduction Sales Company, 60 East 42nd Street, New York 17, has announced that it is marketing two new Heliwelding units; a machine type electrode holder and an automatic unit.

The Heliweld machine holder was designed for mounting on a Radiagraph or other suitable travel mechanism for moving along the joint or for fixed position use with the work moving beneath the arc. It has the same barrel and rack as standard 1⅜-inch diameter machine gas cutting torches.

The Airco automatic Heliweld unit, designed for continuous operation, consists of a holder, holder carriage and a control box. It is fully automatic and electronically controlled. By compensating for set-up irregularities it thereby assures a continuous weld of consistent dimensions and quality.

Both units have a welding current capacity of 300 amperes and are fully

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METALLIC FILTER *Cloth*

Newark Monel and Stainless Steel Filter Cloths are ideally suited to the requirements of refinery filtration. High corrosion resistance, accurate mesh, uniform weave, easy cake discharge and exceptionally long service life are among the advantages offered by Newark filter cloth.

This cloth is available in a wide range of meshes. Let us supply your routine plant and laboratory requirements from stock and let our engineers assist you on any special problems.

Newark Wire Cloth COMPANY

334 VERONA AVENUE • NEWARK 4, N. J.

insulated for the high-frequency current. The machine holder may be used with either A.C. or D.C. welding current while the automatic unit, because of the nature of its electronic circuits, operates on D.C. only.

5—Surface Safety Valve

Otis Pressure Control, Inc., Dallas, has announced a new heavy-duty automatic surface safety valve designed to prevent line breaks caused by excessive pressure increase, or to shut off a broken line.

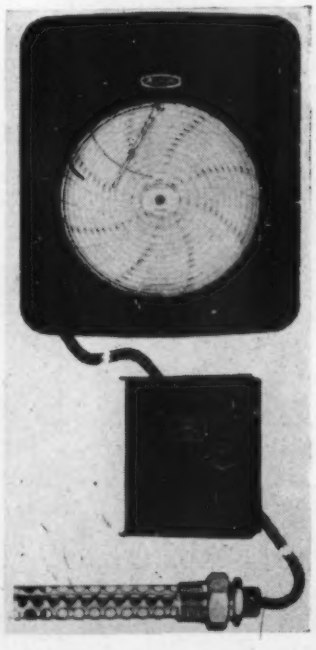
The valve is a simply constructed, self-contained, pressure-operated, full opening gate valve. Operating pressure can be supplied by the line in which the safety valve is installed, or it can be furnished through a connection to an outside source. Operation is identical, regardless of the source of pressure. In either case, no electrical connections or remote control mechanisms are required.

Valve body and gate are of conventional design and are available in several makes, or types, in any standard size and test pressure.

6—Dew Point Recorders

Dew point recorders and recording controllers, said to be the first simple, continuous dew point measuring instruments available for industrial use, are announced by The Foxboro Company, Foxboro, Mass.

The unique feature of the equipment is Foxboro's patented measuring element, the Dewcel. Moisture determination by the Dewcel is based on the fact that for every water vapor pressure in contact with a saturated salt solution there is an equilibrium temperature at which the solution neither absorbs nor gives up moisture to the atmosphere. The function of the variable heat supply is to bring the Dewcel to this equilibrim temperature. This temperature, measured by the thermal bulb, is transmitted to the recording instrument, and is read on the chart in degrees of dew point temperature. Recording is continuous. Various standard chart ranges are



Foxboro Dew Point Recorder

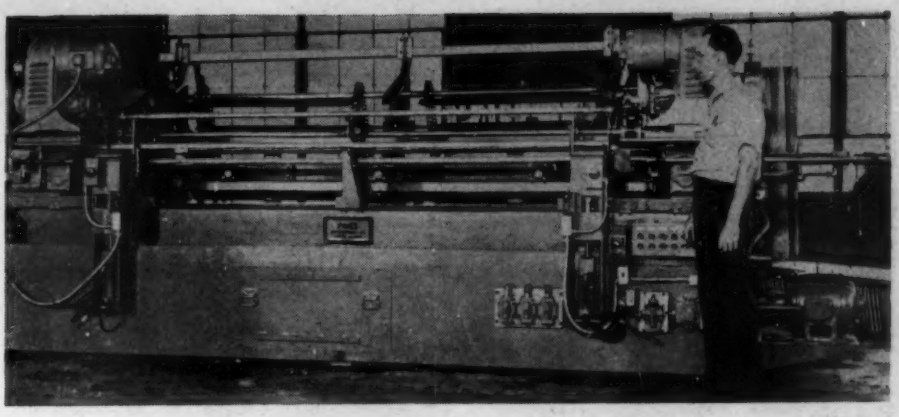
available. The recorders, require no refrigeration, water supply, or air circulating system, and have no mirrors to be kept clean. They can be installed almost anywhere, indoors, or outdoors. Ask for Bulletin 407.

7—Lubricated Plug Valve

H. K. Porter Company, Inc., 1932 Oliver Building, Pittsburgh 22, has announced development of an advanced design of lubricated plug valves, designed to "assure complete and leak-proof shutoff, elimination of leaks through the stuffing box, ease of operation," etc. This new plug valve, particularly applicable to manufactured and natural gas operations, refinery installations, oil production, and the chemical and processing industries, is available in semi-steel, carbon steel, alloy and stainless steels.

8—Profiling Machine

Pines Engineering Company, 1653 Walnut Street, Aurora, Ill., has developed general duty, automatic profiling machines for high production end-machining operations. These machines are used for turning, boring, chamfering, burring, threading, tapping, etc., on the ends of pipe, tubing, bars, fittings, and other work of this type.



Profiling Machine Made by Pines Engineering Company.

Pines profilers are built for either manual of fully automatic operation. Automatic types are equipped with hydraulic controls which repeat the machining sequence continuously. With the work hydraulically chucked, one or both ends are machined by the action of rotating heads, which hold the cutters. One operator can attend several machines.

Either one or two heads are employed. Two heads permit the simultaneous machining of both ends of the work. The profilers may be set up to handle various lengths and diameters of work.

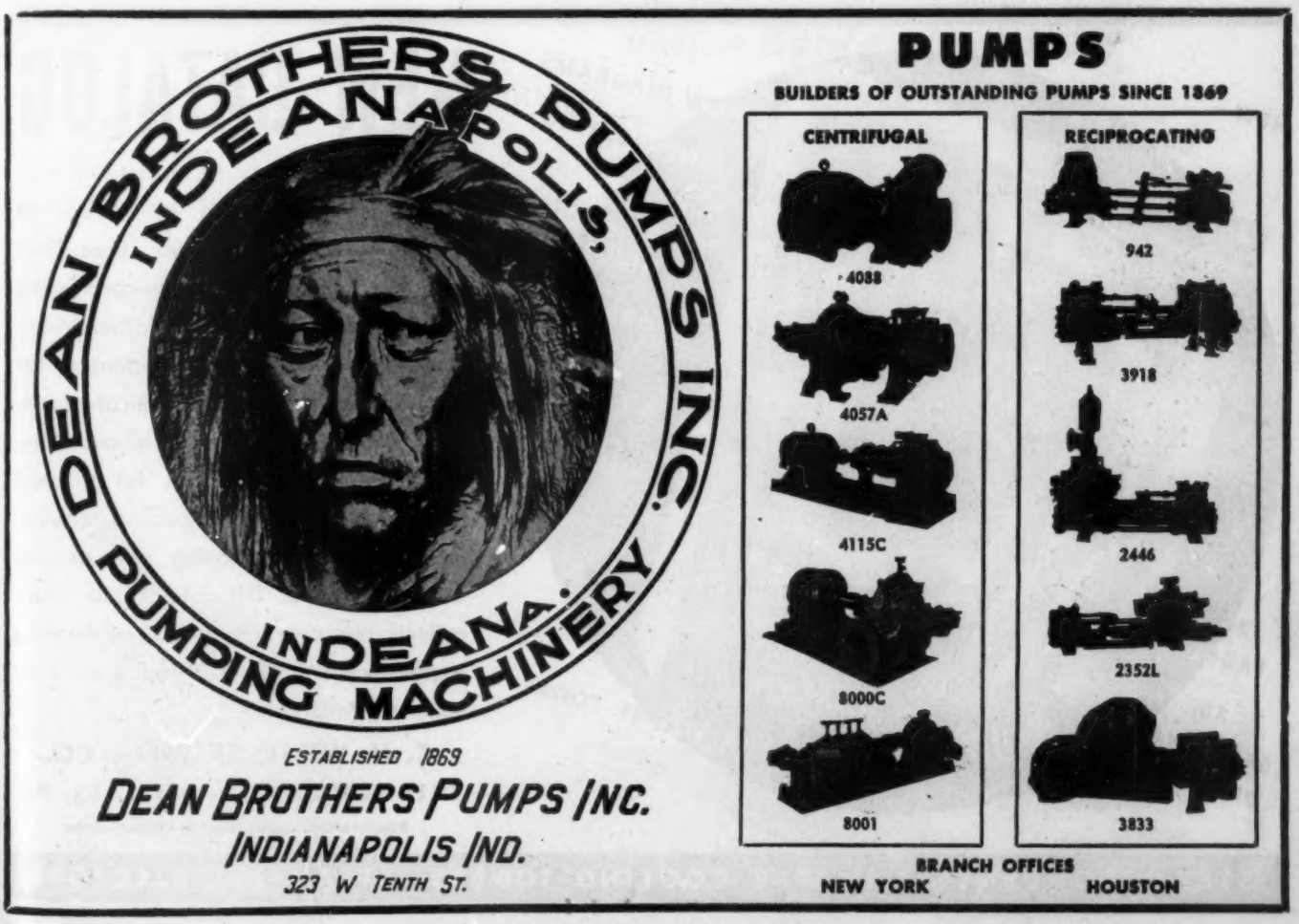
When writing the manufacturer of new equipment for literature or more information identify his product by name and number listed in Petroleum Refiner.

9—Nylon in New Gauge

The Ahscroft "Duragauge," a product of Manning, Maxwell & Moore, Inc., Bridgeport 2, Conn., will shortly be available with a new movement in which duPont nylon will be employed. Nylon, with its low coefficient of friction, high resistance to shock loading, vibration absorbent characteristics, and high corrosion resistant properties, surpasses any material formerly available for this purpose, the company said. This new movement will outwear and outperform any other movement, even under the most severe conditions of pulsation and vibration, the statement continued.

10—Angle-Type Strainer

A strainer with an angle construction so that it may be installed directly upon the unit that it is to protect has been announced by J. A. Zurn Manufacturing Company, Erie, Pa. The strainer thus can be installed when the equipment is



Angle-Type Strainer by Zurn

first installed, or even on the original equipment by the manufacturer, thereby saving time in handling, etc. The strainer also may be used at any point in a pipe line where it can serve as, or replace a right angle connection, thereby saving space and additional connections.

Various types are made with pressure ratings from 125 to 300 pounds. The body and cover assemblies are available in cast steel, semi-steel or bronze.

11—CO_2 Indicator

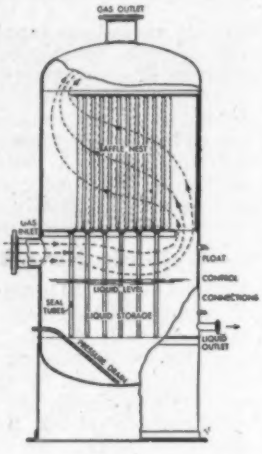
Furnace efficiency is checked with the CO_2 indicator announced by Burrell Technical Supply Company, 1942 Fifth Avenue, Pittsburgh 19. The CO_2 content of the stack is measured in less than a minute by the easily manipulated instrument which can be successfully handled, Burrell says, by an unskilled operator. It is constructed of sturdy, lightweight plastic and metal parts.

12—Thread Compound

Fel-Pro "Hi-Temp" thread compound, that is said to prevent freezing and galling of metal surfaces at temperatures as high as 1800° F., has been announced by Felt Products Mfg. Company, 1504 Carroll Avenue, Chicago. The new C-5 "Hi-Temp" compound was developed specifically for high temperature and high pressure applications, such as furnace tube header studs and plugs. New data sheets, together with a sample of the material are available.

13—"Mist-D-Fier"

A new principle of baffle design for complete mist separation has been developed by J. A. Campbell Company, 645 East Wardlow Road, Long Beach 7, Calif. Known as the "Screenless Mist-D-Fier," this gas separator embodies a series of curved welded baffles installed in staggered rows. Gas passing between these baffles must turn sharply to gain the staggered openings of the next row, and so on, through eleven rows. The mist particles are thrown out of the flow by centrifugal action and impinge on the curved surface of the baffle where they agglomerate into a film. Gas impact causes this film to creep along the curved surface of the baffle, forcing the liquid into vertical slots where it trickles down to the bottom of the baffle into seal tubes and flows down to the bottom trap. The clean gas is directed upward to the outlet at the top of the "Mist-D-Fier."



The device removes all traces of crude oil from natural gas entering an absorption plant; also oil mist and water from gas leaving the plant. A new bulletin suggests other uses.

14—Measuring Instruments

New electric measuring instruments, capable of providing a continuous analysis of flowing streams of fluids and gases, together with automatic control of those processes were described recently by Everett S. Lee, chief engineer of General Electric's general engineering and consulting laboratory, Schenectady 5, N. Y. The X-ray photometer, which he described, provides economical and rapid means of making chemical comparisons, it being possible to determine with it the tetraethyl lead content of gasoline, the concentration of an acid in water, the percent chlorination of a plastic, or the percent ash in coal. This is done by measuring and comparing X-ray absorption of a sample and a reference.

X-ray absorption always is the same for a given material and whether the element is alone or in chemical combination. Analysis is made without loss or alteration of the sample tested.

NEW CATALOG

Theory and Operating Characteristics of Steam Jet Ejectors for all classes of vacuum service. Single, Two, Three, Four and Five Stage Types. Non-condensing and Condensing types with Barometric or Surface Inter- and After-condensers for vacuum requirements in chemical plants, food plants, sugar refineries, oil refineries, power plants, etc. Steam Jet Vacuum Refrigeration for water cooling requirements of air conditioning and process applications. . . . This catalog No. 1462 reflects our over 30 years' pioneering experience. Write for it today on your company letterhead.

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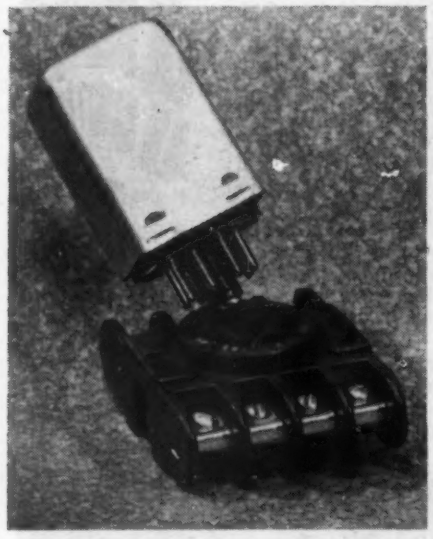
STEAM CONDENSERS—WATER COOLING TOWERS—STEAM JET EJECTORS

Requests for manufacturers' literature get prompt replies when you identify each item by number and name listed in Petroleum Refiner.

15—Relay

An interesting application for a totally-enclosed relay has been reported by the manufacturer, Sigma Instruments, Inc., Boston 21, Mass. An effective means of accurate temperature control consists of a sensitive mercury thermostat which switches the coil circuit of a relay, the relay in turn controlling the load. The Sigma Series 41 relay, suited for this application, has one standard type which operates with only



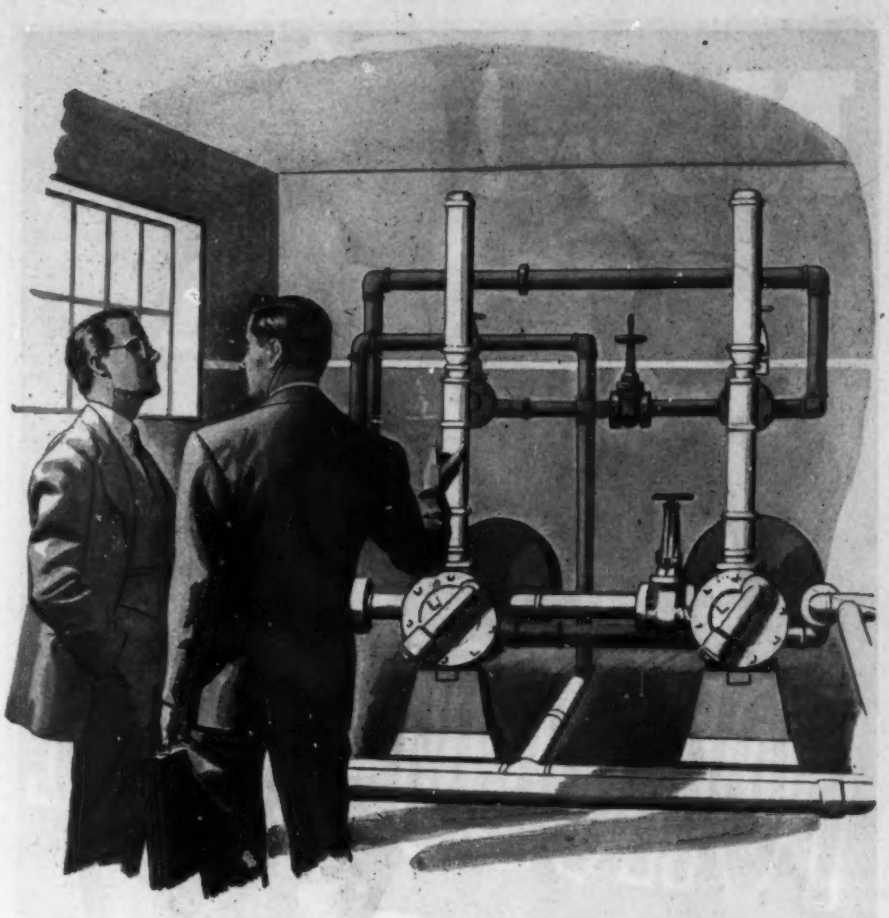
two milliamperes at 115 Volts A.C. in the coil circuit. Replacement after accidental damage or long service is made easy by plug-in mountings.

16—Trip-Proof Coupling

This coupling provides a positive shut-off valve in the same assembly of a Roylyn quick coupling by meshing a valve lever with the locking collar of the coupling. The valve must be in the shut-off position before the coupling can be broken and, conversely, the coupling must be locked in its mating nipple before the valve can be opened.

The piston type valve effects a seal even with static pressure. Made of non-sparking bronze alloy, the ruggedly con-

Roylyn Trip-Proof Coupling



"For fighting fires here, you can't beat a Built-in System!"

Planning to install new fire protection equipment in pump rooms—or other spots where flammable liquids are handled or stored?

You'll find it worth while to talk to a Kidde representative!

He can help you decide just what kind of equipment will most effectively—and economically—safeguard each danger-spot... built-in systems, wheeled units or portable extinguishers.

Whatever type of *Kidde* equipment he recommends, you can be sure of fast, sure action against fires in flammable liquids (in electrical equipment too). The dry, clean carbon dioxide (CO_2) discharged by *Kidde* equipment cannot dilute or contaminate liquids, nor corrode metal parts. No after-fire mess either!

The Kidde representative is always ready to discuss your fire protection problems with you. Why not call him in today?

Walter Kidde & Company, Inc., 1248 Main Street, Belleville 9, N. J.

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Kidde

FIRE PROTECTION HEADQUARTERS

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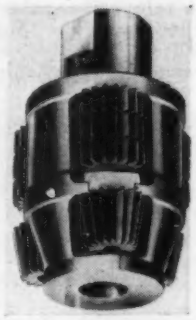
FORGED STEEL
THE LATEST DEVELOPMENT IN
CHIKSAN
BALL-BEARING SWIVEL JOINTS
CHIKSAN has adopted forged steel for all Styles and Sizes to and including 2"... in the following pressure ratings: 1,000 psi, 3,000 psi, 12,000 psi and High Temperature (700 psi, 500° F.). This important development now brings even higher quality to CHIKSAN users everywhere. It is tangible evidence of CHIKSAN's policy for continuous development in the design and manufacture of Ball-Bearing Swivel Joints for all purposes.
CHIKSAN REPRESENTATIVES IN PRINCIPAL CITIES
EXCLUSIVE EXPORT REPRESENTATIVE: CHIKSAN EXPORT CO., Brea, California
Branches: New York 7—Houston 1
CHIKSAN
TRADE MARK
BALL BEARING SWIVEL JOINTS FOR ALL PURPOSES
CHIKSAN COMPANY
BREA, CALIFORNIA
New York 7
Houston 1

structed trip-proof coupling weighs but 11¾ pounds. Available now in 2½-inch pipe thread sizes with larger units contemplated. Pressure tight caps, for closing off the open line, complete the installation. The coupling itself requires only a 40° rotation to activate it. It is easily operated, positive, vibration proof, and requires no tools according to Roylyn, Inc., 718 West Wilson Avenue, Glendale 3, Calif.

17—Expansion Cutter Head

A three-tier single expansion cutter head which employs three cylinder cutters an one cone cutter per row is being manufactured by Airetool Manufacturing Company, 316 South Center Street, Springfield, Ohio. This cutter head is designed for cleaning heavy coke deposits from straight tubes 2¼ to 4¾ inches i.d. It is also available with two or four cylinder cutters per row.



The cutters employed are of the new form type which permits repeated tube cleaning operations without tracking or damaging tube walls.

The head is driven by an Airetool motor which is self-feeding and is fitted with renewable pin plates which permit on-the-job disassembly. Bulletins are available upon request.

Hydraulic Tail Gate Loader

18—Tail Gate Truck Loader

A handy, low-priced hydraulic tail gate loader for trucks has been introduced by The Day Company, 306 West 69th Street, Chicago. The lift platform operates over the full distance from ground to truck floor level, raising and lowering loads up to 1200 pounds. The platform remains level throughout its entire travel arc, and swings up to serve as a tail gate when the truck is on the road. The hydraulic pump is driven from the truck's standard power takeoff and the "package" unit fits most standard 1½-ton trucks.

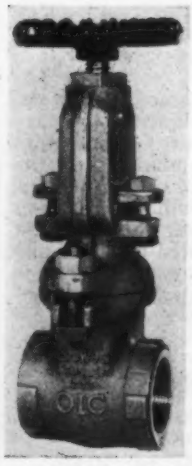
19—Utility Jackhammer

Ingersoll-Rand Company, 11 Broadway, New York 4, has announced its J-10 utility jackhammer, new air-oper-

Ingersoll-Rand J-10 Jackhammer

ated drill, designed for general plant maintenance work. It will do many drilling jobs and its light weight and ease of handling makes it handy for use on ladder, scaffold or for overhead work. It carries a built-in oil reservoir for lubrication. Bulletin 4065.

20—Stainless Steel Valve



A new corrosion-resistant stainless steel valve has joined the line of products of The Ohio Injector Company of Wadsworth, Ohio.

This valve was designed expressly for the chemical, processing, etc., industries, wherever the service calls for corrosion - resistant valves.

The valve, manufactured in ½ to 2-inch sizes, is outside screw and yoke type, having bolted bonnet, tapered wedge, and screwed ends. It is so constructed that all materials in contact with the fluid are of low-carbon type-316 stainless steel. Packing and gaskets are corrosion-resistant. Yoke bushings, through which the stem rises, are renewable.

21—Combination Test Tube

A new combination phosphate mixing and testing tube has been developed by Trutest Laboratories, Inc., 20 S. 33rd Street, Philadelphia 4, for use in water analysis. It eliminates the need for a funnel because the top opening permits filtration of the water sample direct into the tube. This opening also eliminates damage that may be caused by unavoidable spillage of stannous chloride on wearing apparel, etc. This TruTest combination tube fits any TruTest comparator block or Taylor slide.

22—Universal Expansion Joints

Zallea Brothers & Johnson, 888 Locust Street, Wilmington 99, Del., have announced a new line of universal expansion joints especially designed for use in refineries, power plants and chem-

THE SMALL EXTINGUISHER..

WITH THE *MIGHTY PUNCH!*

FOR FIRE HAZARDS THAT DO NOT DEMAND LARGER EQUIPMENT . . .

The new ANSUL 4 Dry Chemical Fire Extinguisher has fire stopping effectiveness far in excess of any other extinguisher of comparable size. Its compact size makes it ideal for locations where space is limited. The ANSUL 4 is easy to use... easy to recharge on-the-spot... and again ready to

STOP FIRE FASTER!

MADE BY THE MANUFACTURERS OF ANSUL-DUGAS DRY CHEMICAL FIRE EXTINGUISHERS, MODELS 20, 30, 150A AND 350A.

ANSUL CHEMICAL COMPANY
FIRE EXTINGUISHER DIVISION, MARINETTE, WISCONSIN

One of two MURRAY multistage steam turbines, designed to deliver 685 horsepower on steam at 110 pounds and exhausting to 26½ inches Hg vacuum, furnished to drive high speed centrifugal compressors for refrigeration in one of the most prominent Southern Hotels. A third unit has recently been shipped. Equipment includes variable speed oil relay governor, trip and throttle valve, strainer surrounding the double seated balanced governor valve, and force feed lubrication.

MURRAY

MULTISTAGE

STEAM TURBINES

★ ★ ★ ★

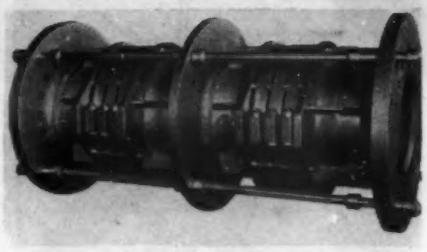
The complete MURRAY line of steam turbines includes all variations of mechanical drives, and turbine generator sets up to and including 3000 KW rating.

Sales Representatives for Murray products are located in all principal cities.

MURRAY IRON WORKS COMPANY

BURLINGTON, IOWA

Builders of Steam Power Equipment for Three Quarters of a Century



Universal Expansion Joints

ical process plants. This Zallea universal expansion joint, capable of absorbing movement of high-temperature lines in any direction—axially, laterally or angularly, or any combination of these, consists of two corrugated elements linked together by a section of pipe.

Particularly suited for catalyst lines —since the joint offers no sharp crevices in which catalyst can pack—the Zallea universal expansion joint has numerous applications in reactor risers and standpipes, regenerator, standpipes, stripper and flue gas lines. These universal expansion joints can be designed for all temperatures and pressures. Ask for Catalog 47.

23—Absolute Pressure Gauge

Development of a new, accurate absolute pressure gauge has been announced by United States Gauge, Division of American Machine and Metals, Inc., Sellersville, Pa.

The new gauge is designed to indicate, in units of inches or millimeters of mercury, the absolute pressure or vacuum to be measured. The graduations on the dial are the same as used on barometers.

This gauge can be built in a number of ranges to suit specific requirements. The case, of black molded phenolic material, permits the gauge to be used in highly contaminated or corrosive atmospheres.

24—Weatherproofing Material

A new tough but elastic coating material for "weatherproofing" outdoor metal surfaces has been announced by The Arco Company, Cleveland (7301 Beesemer Avenue) and Los Angeles.

This "Dum Dum for Metal" provides a durable coating many times the thickness of paint that makes the metal surface to which applied impervious to measurable vapor and moisture transmission, acid and alkali gases and fumes and other similar causes of deterioration, thereby preventing surface corrosion and disintegration due to these conditions, the manufacturers say. It is applied by spray method. The material is not recommended for use over bituminous-type coatings.

25—Packing

A new type of packing, specially developed for the petroleum industry, has been announced by Quaker Rubber Corporation, Tacony and Comly Streets, Philadelphia 24. This quality packing (Petro-pak No. 1675) is ideal for tight-sealing service on refinery and field equipment handling oils, gasoline, and other volatiles at normal atmospheric temperature.

"Petro-pak" is made from a special rayon roving, developed mainly for

mechanical packings. This is saturated with a special lubricant that will not dissolve or wash out even when packing is subjected to the effects of oil, gasoline or similar solvents. The packing is available in rolls, in sizes from ¼ to 1 inch, inclusive, in units of 1/16-inch.

For free copies of manufacturers' literature or more information about new equipment featured in Petroleum Refiner, simply identify the product by name and number.

26—Tube Fitting

A new type tube fitting employing a hardened alloy spring steel sleeve is announced by the Flodar Corporation, 331 Frankfort Avenue, Cleveland, Ohio. The fitting consists of three parts; a connector body, a tightening nut, and the contractible sleeve which grips the tube with slotted, spring steel fingers when the tightening nut is turned.

Flodar grip tube fittings are manufactured in sizes ¼ to 1½-inches and are available in straight, union, elbow, side tee, tube tee, and cross, both male and female.

27—Connectors for Grounding

"Connectors for Grounding" (Catalog G 47) published by Burndy Engineering Company, 107 Breckner Boulevard, New York 54, presents the company's line of grounding devices and in addition, provides a comprehensive discussion of modern grounding practice.

28—Refractory Backing

"The most advanced wall design in years" is the claim made by The George P. Reintjes Company, 2517 Jefferson Street, Kansas City 10, Mo., for its flexible, unit supported, insulated refractory backing for close-spaced water walls. Individual tiles are used between the tubes, the tiles can move with the tubes in any direction, it is lightweight, there is no over-heating of stud bolts, no wire mesh or expanded metal is utilized near the furnace wall, etc. Heat transfer through is approximately one third the heat transfer through dense refractory. Ask for Bulletin 947.

29—"Pancake Motor"

Fairbanks, Morse & Company, 600 South Michigan Avenue, Chicago 5, have issued Bulletin AEB210.1 covering its axial air gap motor, sometimes called the "pancake motor." This motor is distinguished by the fact that the air gap dimension is measured parallel to the axis of the shaft, as compared to a radial dimension in the regular type. It is not intended to supersede or replace the conventional type motor and the bulletin outlines the new motor's advantages for certain definite applications. Features of construction, its versatility, etc. are shown.

30—"Wetter Water"

"Penetrate," a new American-La France-Foamite Corporation (Elmira, N. Y.) product compounded of special ingredients, "1 percent of which when added to ordinary water gives the combined solution speedy penetrating and spreading qualities much superior to water itself and therefore, more effective in fire fighting." Ask for Bulletin AD 9010.

31—Valves and Controllers

Data Unit No. S-208, has been issued by The Swartwout Company, 18511 Euclid Avenue, Cleveland 12, Ohio, describing its new V10 flow regulating valve. Design details are described and pictured to reveal features appropriate for use in handling liquids at high pressure drop in connection with drainage and flashing condensate control.

Bulletin S-17 illustrates and explains a new series of precision differential pressure master controls.

32—Meters

Bulletin 32-F announcing a complete line of corrosive service flow meters has been issued by Fischer & Porter Company, Hatboro, Pa. Described are instruments made of various corrosion resistant materials and also included are dimensional details and a section on laboratory type flow meters.

33—V-Belt Drives

A new 144-page, indexed catalog for "pre-engineered: Texrope" drives which simplifies drive selection so persons untrained in V-belt engineering can readily specify proper drives, has been announced by the Allis-Chalmers Manufacturing Company, Box 512, Milwaukee, Wis. The catalog is said to represent one of the most outstanding compilations of engineering data ever assembled for the V-belt industry.

Nearly two-thirds of the book is devoted to pre-engineered stock Texrope

CAST IRON & CAST STEEL
WEDGE GATE VALVES
A.S.A. SERIES 125-1500
This important range of valves for the Oil Industry are available as follows:—Cast Iron Series 125, Cast Steel Series 150, 300, 400, 600, 900 and 1,500. They conform to API Std. 600A as regards design, dimensions and materials. Illustration shows our No. 581 Series 300. Fully descriptive literature available on request.
ONE OF THE WIDE RANGE OF
OUR SPECIALISED PRODUCTIONS
FOR THE OIL INDUSTRY
Valves
Specify them for...
OIL LINES &
REFINERY
SERVICES
Newman, Hender & Co. Ltd
WOODCHESTER GLOS. ENGLAND

drives for all applications from one to 150 horsepower. More than 22,000 stock drives are listed. Drives for all horsepowers, motor speeds, ratios and driven speeds have been accurately pre-engineered and are shown.

34—Self-Priming Pump

Allis-Chalmers Manufacturing Company (Milwaukee 1, Wis.) automatic, self-priming pump equipped with a new type automatic spring valve which gives "faster, smoother transition from priming or vacuum pumping to straight centrifugal action," is described in Bulletin 08B6319B. The pump's portability, rapid installation, quick priming features, and non-clog design of its open impeller types, are portrayed.

35—Floats

Chicago Float Works, Inc., 2330 South Western Avenue, Chicago 8, announces its catalog is available for general distribution for the first time. Detailed are the complete line of copper, steel, monel, and stainless steel industrial floats. The bulletin includes engineering data necessary to make volume and buoyancy calculations, together with diagrams of standard float connections used with float valves and liquid level controllers.

36—Heavy-Oil Burners

A new heavy-oil burner of the horizontal rotary type, with a capacity of 5 to 150 gallons per hour, has been announced by Preferred Utilities Manufacturing Corporation, 1860 Broadway, New York 23. Among its features are the patented "voluvalve," the precision air-oil control, and the "sure-fire" ignition system.

Automatic throttle control assures the maintenance of an accurate combustion mixture throughout the entire range of high-low operation, regardless of the pressure, viscosity and temperature of the oil. Ask for Bulletin 175.

Requests for manufacturers' literature get prompt replies when you identify each item by number and name listed in Petroleum Refiner.

37—Hydraulic Accumulator

A new booklet on hydraulic accumulators is announced by Greer Hydraulics, Inc, 454 18th Street, Brooklyn, 15, N. Y. Included are a history and development of the accumulator, an outline of the main features of the Greer unit,, construction details, materials, principles of operation, performance charts, etc. Typical applications also are shown.

38—Corrosion Resistant Alloy

A "Guide to Selection of Durco Corrosion Resisting Alloys," Bulletin 100, recently published by The Duriron Co., Inc., of Dayton 1, Ohio, lists 155 corrosives in alphabetical order and shows which of five Durco alloys offers most satisfactory resistance to each corrosive.

Where two or more alloys are satisfactory for a particular service, the bulletin enables the user to make a selection on the basis of the most economical alloy. A table shows the mechanical properties of each alloy.

39—Built-in Motors

The Louis Allis Company, Milwaukee 7, Wis., Bulletin 516-D, describes the line of type "CT" rolled shell, shaftless squirrel cage induction motors for built-in drives with details of electrical and mechanical features, mounting, ventilation and applications.

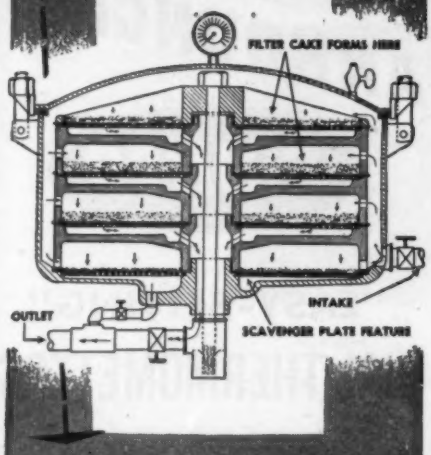
40—Flow Meter

Illustrated Bulletin 35-83A on the Adsco flow meter of the orifice type for indicating, recording and integrating the flow of steam, gas and air gives details of construction, data on orifice plates, etc., may be obtained from American District Steam Company, North Tonawanda, N. Y.

41—Steam Cleaner

A new "utility" model "Hypressure Jenny" steam cleaner announced by Homestead Valve Manufacturing Company, Coraopolis, Pa., is said to be a full-powered, extra heavy duty, all-purpose steam cleaner. The new unit features instant starting simplicity of design and operation; rugged welded unit construction; complete accessibility of all working parts; continuous electric spark ignition; non-clogging heating coil; and pressure atomizing oil burner. Ask for Bulletin 17-A-16.

the horizontal plate principle



. . . makes these the "filters that filter all liquids"—

SPARKLER Horizontal Plate FILTERS

. . . and here's the story:

- Sparkler Horizontal Plate Filters handle any liquid from heavy varnishes to light alcohols, including products such as acids, oils, beverages, milk, pharmaceuticals, plating solutions, water, etc.
- Equally efficient performance on intermittent or continuous operation, under a wide range of temperature, pressure and viscosity conditions.
- Equally effective whether removing carbons and contact clays or clarifying and polishing with filter aids.
- Patented Scavenger Plate permits complete batch filtration. (It's virtually an auxiliary filter with an independent control valve.)
- Unexcelled filter cake stability—no slipping or breaking.

. . . why?

Because filter media are supported on a horizontal plane and filter aids floated into position uniformly, filtration takes place uniformly over entire filtering areas. Flow through filter is always *with* gravity.
Sparkler Filtration Is *Engineered* Filtration—we invite correspondence on your problem. You will receive the advice of filtration scientists with a quarter of a century of experience in a specific field.

SPARKLER MANUFACTURING CO.
MUNDELEIN, ILLINOIS

42—Worm Gearing

The new De Laval worm gearing bulletin describes advantages of worm gearing, AGMA service classification, bearing load formulae, quality control and precision manufacturing, selection practice, horsepower ratings, worm thread and gear tooth data for different ratios, dimensions and details of standard worm mountings, worm dimensions, gear dimensions and dimensions of standard flanged rims. Bulletin WG1220-5-46, De Laval Steam Turbine Company, Trenton 2, N. J.

43—Steam Purifiers Blowoff Systems

Publication 4410 by Cochrane Corporation, 17th and Allegheny, Philadelphia 32, describes five different flash-tank and heat-exchanger systems for recovering the otherwise wasted heat in boiler blowoff water. Detailed calculation of savings under actual plant conditions and a quick-estimating chart are provided. Two thermostatically controlled systems are described for automatically proportioning the rate of blowoff to makeup water.

The same company's Publication 2725 describes steam purification units for removal of moisture and solids in steam lines.

44—Copper Sweetening Process

Air Reduction Company, 60 East 42nd Street, New York 17, has published a folder, complete with diagrams, which explains the use and application of the Airco-Hoover copper sweetening process. This process is the original method of "sweetening" malodorous oils by bringing them in contact with copper chloride.

45—Metal Tubing Circuits

An authoritative handbook on the selection, sizing, layout and installation of metal tubing circuits has been published by The Parker Appliance Company, 17325 Euclid Avenue, Cleveland, Ohio. Called the "Tube Fitter's Manual," the handbook makes available basic data underlying the specialized technique required for the design and speedy installation of long-life, leakproof systems.

Available on request, the manual should be particularly useful to personnel of design, assembly and plant engineering departments of companies throughout industry generally.

46—Turbo Pumps

Bulletins by The J. S. Coffin, Jr., Company, 326 South Dean Street, Englewood, N. J., describes in detail that company's high speed, single stage turbo pumps, portraying with schematic drawings and photographs the several models and giving ratings, etc.

47—Welding Holder and Head

Air Reduction Sales Company, 60 East 42nd street, New York 17, has published a folder describing its three new types of Heliweld equipment. Including the manual holder, the machine holder, and the electronically controlled automatic head. The folder lists advantages and uses of the process, construc-

. but the designing of a in calls for the services of ery few skills, whereas the designing of an oil re-ery or a chemical plant mands a FORMIDABLE RAY OF SKILLED MAN-WER.

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Even large cracks which you might think would require replacement of the unit, can be repaired with Smooth-On. Most such Smooth-On repairs can be made without dismantling . . . and never require heat.

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FREE REPAIR HANDBOOK

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Do it with SMOOTH-ON
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tion details of the equipment and a page devoted to supplementary equipment and supplies. Write Department 1632.

48—Pumping, Heating Units

Fuel oil pumping and heating units for preparing for combustion heavy fuel oil for all types of oil burners, are described in Bulletin 40, National Airoil Burner Company, 1284 East Sedgley Avenue, Philadelphia 34. Various applications of the several types are depicted with photographs and schematic art and ratings and specifications are set forth in great detail. Useful data for general use by engineers are included.

49—Bent Tube Boiler

Henry Vogt Machine Company, 10th and Ormsby Streets, Louisville, Ky., has issued a new bulletin covering its Class LE water tube boiler, "best adapted to installations where there are no space limitations." Elaborate sketches of a score of previous installations are presented together with photographs of actual manufacturing processes. Details of furnace water wall construction are shown for four typical designs of the Vogt water-cooled wall construction.

50—Gas Mixers

Bulletin 5C-1 has been issued by Bryant Heater Company, 1020 London Road, Cleveland 10, Ohio, detailing its "hijectors, Venturi mixers for high-pressure gases" which are designed to entrain all the air needed to make a complete combustible mixture and to deliver it at the highest possible pressure. Where high-pressure gas is available, "hijectors" eliminate the requirement of air blowers or pressure aid for combustion jobs. They are easy to install and operate, Bryant claims.

51—Protective Coating

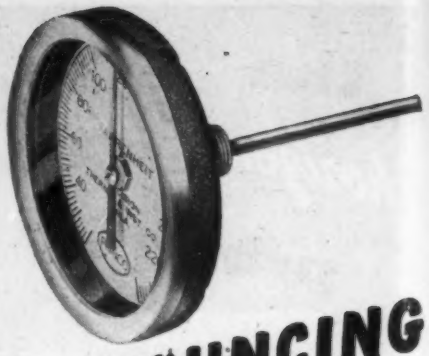
A catalog giving details on the chemical and physical properties, application instructions and other data on protective coatings is available upon inquiry to the Amercoat Division, American Pipe and Construction Company, P. O. Box 3428, Terminal Annex, Los Angeles 54. Catalogs are available for 10 major industries.

52—Fuel Testing

New mailing pieces describing "Knockometers" for automotive aviation fuel testing on single cylinder laboratory engines have been issued by Sperry Gyroscope Company, Inc., Great Neck, N. Y. Described are the operation and the specifications of "Knockometer" models K-1 (aviation) and K-3 (automotive) under Bulletins 15-46 and 15-47, respectively.

53—Oil Hose

A new catalog section on its oil hose for cargo handling has been published by The B. F. Goodrich Company, Akron, Ohio. This combination suction and discharge hose is one of the most highly specialized products of the rubber industry. Types described include dock loading, sea loading and hose for barge service, as well as the fittings.



ANNOUNCING

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"EASY-SETTING"
DIAL THERMOMETERS

- 5 Standard Ranges — Fahrenheit or Centigrade
- "Easy-Access" Cases for Quick Setting
- Simplified Dial and Pointer Adjustment
- Rugged, Accurate, Bimetallic Unit

"Easy-Access" cases permit all adjustments to be made *from the front,* "on the job"!

Two case types: "Quick-Opening," with *threaded frame;* "Moisture-Proof," with *bolted frame.*

4" dial. 5 standard ranges, Fahrenheit or equivalent Centigrade, covering —100° to +850° F., or —75° to +450° C.

Stainless Steel stem in 4 standard lengths: 4", 6", 9" and 12". Separable sockets available. *Write for illustrated folder.*

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About EQUIPMENT—SERVICE Suppliers

Kerotest Appoints Isaacks To Works Manager Position

J. D. Isaacs has been named works manager of Kerotest Manufacturing Company, Pittsburgh. A greatly expanded program in the production of Kerotest pipe line and refinery valves to meet current oil country developments is linked with the appointment of Isaacks who has been Southwest division manager for Kerotest headquartering in Houston. Isaacs has been associated with Kerotest Manufacturing Company for more than 14 years.

Isaacks

Named by Hammel-Dahl

The Hammel-Dahl Company, Providence, R. I., has appointed Metrol Company, Detroit, Mich., as sales representative in the lower peninsula of Michigan and Northwest Ohio.

Metrol has for a number of years specialized in the sales and engineering of automatic control and instruments in the industrial process and power field and is headed by R. A. Wright, president.

Oceco Appoints New Managers In New York and Pittsburgh

The Oceco Division of The Johnston & Jennings Company has announced the following changes effective December 1:

J. R. Edwards, Sr., formerly Pittsburgh manager, to New York where he will succeed George T. Bloom as district manager. Bloom is retiring after 25 years of service.

J. R. Edwards, Jr., formerly assistant Pittsburgh manager, advanced to the manager's position.

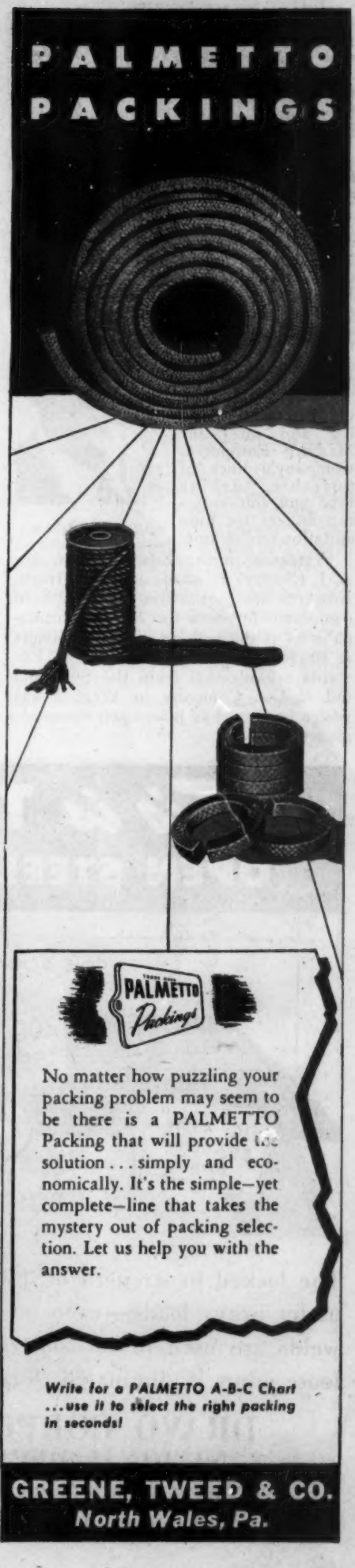
Del Heideman, development engineer at Cleveland, to New York as office engineer.

Refinery Supply Planning Modern Addition at Tulsa

Refinery Supply Company, Tulsa, has announced that it will construct a new building to house the manufacturing and storage facilities of the company. Latest type of equipment needed to make, repair and assemble materials used in refineries and gasoline plants will be embodied in the plans, W. A. Schlueter, founder and president, announced. The new building, modernistic in every re-

Oakite Representatives in Sectional Meetings

Technical sales conferences were held in the fall by Oakite Products, Inc., manufacturers of industrial cleaning and related materials and at which were featured special reports from the company's research chemists, service engineers and other specialists. The top group above shows members of the Midwest, Southwestern and Southern divisions gathered in St. Louis and the lower group was snapped in Chicago where the Central, Chicago and Detroit divisions met.


PALMETTO
PACKINGS
PALMETTO
Packings
No matter how puzzling your packing problem may seem to be there is a PALMETTO Packing that will provide the solution . . . simply and economically. It's the simple—yet complete—line that takes the mystery out of packing selection. Let us help you with the answer.
Write for a PALMETTO A-B-C Chart ...use it to select the right packing in seconds!
GREENE, TWEED & CO.
North Wales, Pa.

spect, will adjoin present facilities of the company and will be located on a lot now used for parking.

Schlueter said that the company now has a backlog of orders "from five to ten months" and the additional room is badly needed.

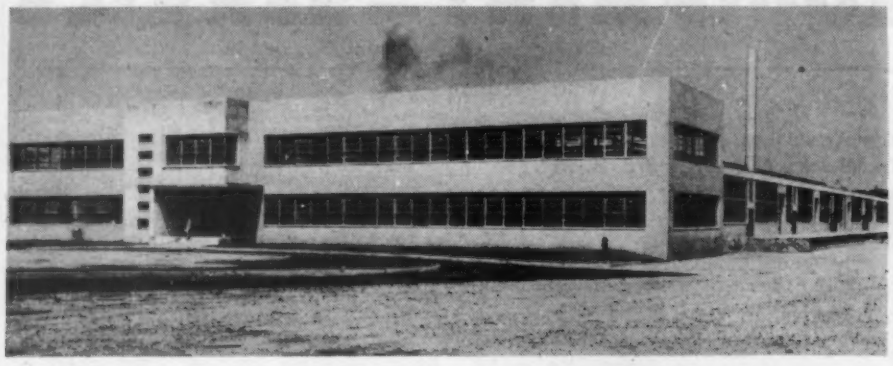
Mixing Equipment Appoints Southwestern Representative

Allan Edwards, Inc., Tulsa, has been appointed exclusive representative by Mixing Equipment Company, Inc., of Rochester, N. Y., for Oklahoma, Kansas and the Texas Panhandle districts including Fort Worth and Dallas territory. The organization will handle engineering and sales for Mixing Equipment Company's lines of portable, turbine type and side entering mixers for fluid agitation and mixing.

Edwards

Principals in the Edwards firm are A. J. (Scotty) Edwards and J. D. Irons. Edwards has been active in the oil field equipment business for 25 years principally as sales manager of Black, Sivalls & Bryson, Inc. Irons came to the Edwards organization from the Stanolind Oil & Gas Company in West Texas where he served as junior petroleum engineer.

Manning, Maxwell & Moore's New Factory



A long term lease on a new factory building in Stratford, Conn., has been signed by Manning, Maxwell & Moore, Inc. This is in addition to their five other plants. Production of the company's Ashcroft gauge line, now manufactured at the Bridgeport plant, will be moved to Stratford, opening the way for expansion of the Consolidated safety valve and American industrial instrument lines at the present Bridgeport location.

Fifth Field Organization Announced by Allis-Chalmers

A fifth region for the field organization of the Allis-Chalmers general machinery division has been announced with Arch J. Cooper, New York district office manager, in charge. Formation of the first four regions was revealed earlier this year. The new area will embrace territory now covered by New York, Buffalo, Rochester, Syracuse and Newark offices.

Cooper, who has been associated with Allis-Chalmers since 1909, continues as New York district office manager in addition to regional manager. Vernon L. Spinney has been named assistant district office manager. William J. Devers has become Newark district office manager, while Ned W. Landis also becomes district manager at Syracuse. H. E. Weiss and Robert G. Burns continue as district manager of the Buffalo office and as special representative of the Rochester office, respectively.

Honeywell Announces Warmee As Moduflow Sales Manager

Roy H. Warmee has been appointed sales manager of the Moduflow division, Minneapolis-Honeywell Regulator Company. He has been sales promotion manager since 1940 and with the company since 1925.

Warmee is regional director of the National Federation of Sales Executives and president of the Minneapolis Association of Sales Managers. A writer and lecturer on sales subjects, he was the first winner of the Howard G. Ford award for outstanding achievement in sales management. He also has won the Edmand F. Mair trophy, for "outstanding service to Minneapolis sales management."

John Randall, until recently temporary manager of the Moduflow division, has joined another company. Sales promotion activities will henceforth be managed by John A. Young, who has been serving as Warmee's assistant.

Oil Well Representative

Oil Well Supply Company has named Donald F. Greenlaw to serve the Corpus Christi, Texas, area. Greenlaw, who has been transferred from San Antonio to Corpus Christi, was a captain in World War II. He joined Oilwell late in 1945, and served in stores at Alice and Pettus, Texas, until January 1, 1947, when he transferred to the company's Oil City plant for a student engineering training course. He completed this course in July.

Wolverine Opens Warehouse

Wolverine Tube Division's first postwar warehouse was opened recently at 1515 Fulton Street in Houston. The warehouse will stock Wolverine seamless non-ferrous tubing in the form of condenser tube, S.P.S. pipe, copper water tube and dehydrated refrigeration tube.

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The locked in strength of TRI-LOK enables it to stand up under heavy loads—even on long spans. No rivets, bolts or welds are used in its construction, thus, the possibility of loose joints is eliminated. *Write for Bulletin 1140.*

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(Distributor for THE TRI-LOK COMPANY)

DRAVO

Electro-Alloys Division Opens Modern New Foundry

"The most modern foundry in the country today" was shown to hundreds of special visitors who in October attended the formal opening at Elyria, Ohio, of the new foundry of Electro-Alloys Division of American Brake Shoe Company. This plant, the result of years of research by Brake Shoe's engineers and metallurgists, combines the finest equipment, layout and casting methods known to the high alloy casting industry, officials declared.

Electro-alloys radiographers marking up a muffle section for X-ray examination

One of the many modern facilities which the visitors saw was the X-ray inspection of castings. All castings for certain specific operating conditions are subject to thorough X-ray examinations. This includes muffles, pots, retorts, valves and other castings of similar type. **The accompanying picture shows the larger of two units used at Electro-Alloys.** All examinations are conducted by highly trained industrial radiographers who are supervised by the plant metallurgist. The X-ray units are used to disclose internal defects such as shrinks, spongy metal, blow holes, etc., which ordinarily pass other inspection tests.

This foundry, with a capacity of 200 tons of castings a month, "is Brake Shoe's contribution to the solution of one of industry's most costly problems," it was stated.

Mexico Service Established

Chemicals for industrial water conditioning and fuel oil treatment are now available in Mexico from The Bird-Archer Company through their technical representatives, Calderas Y Accessorios, Amsterdam 291, Mexico D. F.

Angel Noriega, general manager of the Mexican firm, has announced that Calderas Y Accessorios have established a complete technical service in Mexico to facilitate the sale and service of Bird-Archer chemicals.

Foote Bros. Corporation Promote Moir, Quackenbush

Foote Bros. Gear and Machine Corporation last month advanced R. B. Moir to assistant vice president in charge of engineering and product development of the industrial gear division. Moir is a

Quackenbush

Moir

graduate in mechanical engineering at the University of Illinois.

B. H. Quackenbush, another M.E. graduate from Illinois, formerly assistant sales manager, will assume the full responsibility of sales manager of this division. Quackenbush has been associated with Foote Bros. for 14 years.

Celanese Transfers Schnabel

William H. Schnabel, superintendent since 1944 of the acetic acid unit at the Bishop, Texas, plant of Celanese Corporation of America, has been transferred to New York as director of the new technical service department of Celanese Chemical Corporation. This new department has been established to

Modern petroleum processing methods

Here is a guidebook for plant managers, superintendents, chemists, engineers, and other refinery men—providing a complete, modern treatment of petroleum processing. It emphasizes the relation of refining to chemical engineering, covering fundamentals, many details of the design and operation of petroleum equipment, economical aspects, etc.

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It provides a complete background of engineering theory in chapters on chemistry, laboratory evaluation methods, the physical properties of oil, and the unit operations of chemical engineering as applied to petroleum—as well as chapters that deal particularly with petroleum refining.

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Fig. 21 Fig. 22

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Fig. 4-F

Fig. 13

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coordinate production quality control with customer needs, to provide customers with technical knowledge in production problems, and to assist in the development of improved methods of handling the production and control of chemicals, it was explained.

Schnabel has been with Celanese Corporation of America since 1940.

Nowery Smith Company Selected as Distributors

Nowery J. Smith recently resigned as vice president of Parkersburg Rig & Reel Company to devote full attention to Nowery J. Smith Company which is specializing in the corrosion field as commercial hot-dip galvanizers, electroplating, pickling and oiling of mill steel products.

Smith

Nowery J. Smith Company recently was appointed Southwest distributors and applicators for Amercoat corrosion resistant Vinyl and Thermo-setting plastic coatings. The company also is distributor of Amercoat plastics to the petroleum industry in foreign fields.

New Divisional Sales Head

Gerald J. Klopfenstein has been appointed sales manager, meter division of Bowser, Inc., Fort Wayne, Ind. Klopfenstein joined the organization in October 1923, and has served in many departments. Prior to his new assignment he was manager of planning and sales engineering.

Product Name Changed

Pennsylvania Salt Manufacturing Company's product formerly called Asplit F cement has been changed to Pennsalt HF cement. It was explained the new name is more in conformity with other company named products, with the "HF" denoting the cement's complete resistance to hydrofluoric acid in all strengths.

Mannion Is Vice President Of Trinity Equipment Corp.

Trinity Equipment Corporation of Orange and Newark, N. J., has appointed James V. Mannion vice president in charge of manufacturing.

Mannion

Mannion joined Brown Instrument Company in the late twenties and after several years with that firm in the Mid-Continent, he located with Phillips Petroleum Company at their Kansas City refinery. He came to New York to join the engineering staff at Alco Products and shortly before the war left there to take over the instrument department of Foster Wheeler Corporation oil division.

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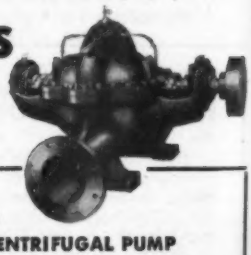

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Hewitt Names Representative

Industrial Supply Company of Minneapolis, has been named distributor of the entire Hewitt Rubber Division's line of industrial hose, belting and packing. Industrial thus will serve all industries with the Hewitt line covering Minneapolis, St. Paul, Minnesota, and North and South Dakota.

Industrial officers include E. F. Rampfer, president, and V. W. Olson, vice president.

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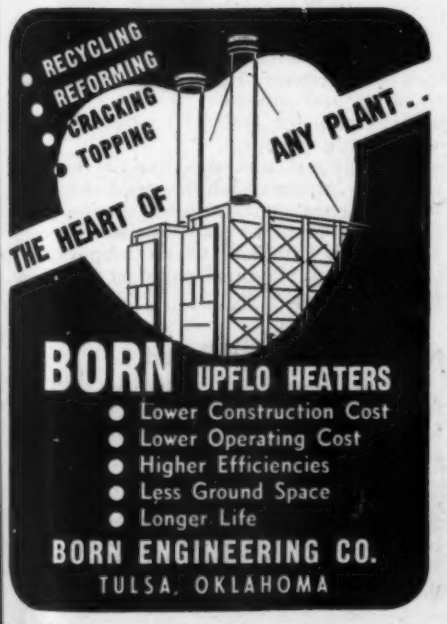

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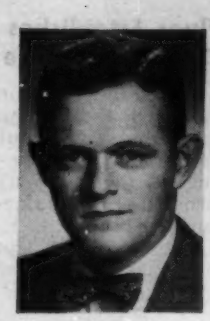

Tyler

Two Promotions Announced By Tube Turns, Inc.

Two promotions have been announced by the welding fittings division of Tube Turns, Inc., Louisville, Ky. Norton P. Bosemer has been placed in charge of the Los Angeles office, 1489 Washington Boulevard, and Robert S. Tyler, Jr. has been placed in charge of a newly established Mid-Continent district, 311 Tuloma Building, Tulsa.

Bosemer joined Tube Turns sales staff in April, 1937, and was assigned to the San Francisco office in May, 1944. Tyler became associated with Tube Turns, Inc. in January, 1946, after four years of service with the Navy, and is a graduate of Tube Turns' sales training project.

Drum Shop Expansion by Babcock & Wilcox Company

An addition to the No. 3 drum shop of The Babcock & Wilcox Company was due for early completion, company officials said last month. This part of the previously announced expansion program, an addition, 275 x 100 feet, will bring the total length of the shop to about 600 feet.

A 100-ton crane will be installed, along with additional welding equipment and drum turners. The new equipment will facilitate handling of the largest drums and vessels fabricated by the company, such as boiler drums for large central stations, pressure vessels and alloy lined vessels for the oil, paper and chemical industries.

Wetzel and Thomas Promoted By Natural Gas Equipment

Natural Gas Equipment Incorporated has assigned A. E. Wetzel to sales, R. H. Thomas succeeding him as office manager. Wetzel has been office manager for NGE for 15 years whereas Thomas has been with the company since 1940. Thomas will be assisted in his new work by Malcolm Parker, former employe of The Texas Company.

The home office of NGE is in Petroleum Building, Los Angeles with branch offices in San Francisco and Bakersfield.

Birthday Celebration

Durametallic Corporation, Kalamazoo, Mich., wished itself a happy birthday in October with a special issue of "The Stuffing Box" wherein policies and accomplishments of the company, which is 30 years old, were reviewed. Featured were photographs on the cover of the "humble birthplace" of Durametallic and alongside its modern building.

Sam Schaeffer, president, in a letter, pays tribute to his associates and thanks customers responsible for the company's growth.

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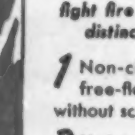

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Fluor Establishes New Customer Service Department

The Fluor Corporation, Ltd., has announced establishment of a customer service department to handle the servicing of their various products of manufactured equipment. E. L. Bishop, who has been employed by Fluor for 12 years in the manufacturing department, will be in charge of this service for the Mid-Continent area, with Houston headquarters. Establishment of such a department to take care of service work, start-ups and any operating troubles which might be encountered, will enable Fluor to render a more complete service to their various clients, officials declared.

Bishop

Sperry Products to Relocate In Dansbury, Connecticut

Sperry Products, Inc., will move its manufacturing plant and general offices from Hoboken, N. J., to Danbury, Conn., during the summer of 1948. A 20-acre site has been acquired and construction of a monitor-type manufacturing building and a three-story office building will begin immediately.

In addition to the testing instruments, Sperry also will manufacture rail detector cars for testing rails in track and hydraulic and mechanical equipment at its new location. Present branch offices in Chicago and St. Louis will be be maintained.

Duenner Opens in Houston To Handle Carlson Stainless

G. O. Carlson, Inc., Thorndale, Pa., has appointed Robert Duenner & Company, 1318 Conti Street, Houston, sales agents for Carlson stainless steel products in the state of Texas. The Duenner company has represented Carlson in Tulsa for some time and the Houston office recently was opened for the express purpose of handling the Carlson line.

John G. McCain, who is in charge of the Houston office, was with Humble Oil & Refining Company, Baytown, Texas, before entering World War II.

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Two Join Powell

The Wm. Powell Company, valve manufacturers of Cincinnati, have announced that S. H. Blackwood, formerly with Mid West Piping & Supply Company, Inc., is now with Powell's New York office at 50 Church Street.

T. J. Malone, formerly with Crane Company, also is now in the New York office of The Wm. Powell Company.

Weeks Given Promotion

Appointment of Charles E. Weeks as assistant to the operating vice president of Southern Alkali Corporation, Corpus Christi, Texas announced late last month by O. N. Stevens, Operating Vice President for the firm.

A native of Runge, Texas, Weeks has been associated with Southern Alkali since the firm was established here in 1934. Prior to this appointment, Weeks had served as auditor for the alkali producing firm. His headquarters will be at the Southern Alkali plant, Corpus Christi.

Weeks

Jebens Joins Patterson

Raymond H. Jebens has joined Patterson Foundry & Machine Company, East Liverpool, Ohio, as chief chemical engineer. Jebens was formerly associated with the DuPont Company where his work included synthesis experimentation and process engineering in organics, heavy chemicals, and paints. He has had long experience in research and design of heat transfer and mixing and processing equipment, and was one of the engineering group which designed the Oak Ridge, Tenn., and Hanford, Wash., plutonium plants. He also has been associated with Kimberly-Clark Corporation since 1945 as chemical engineer in process development work.

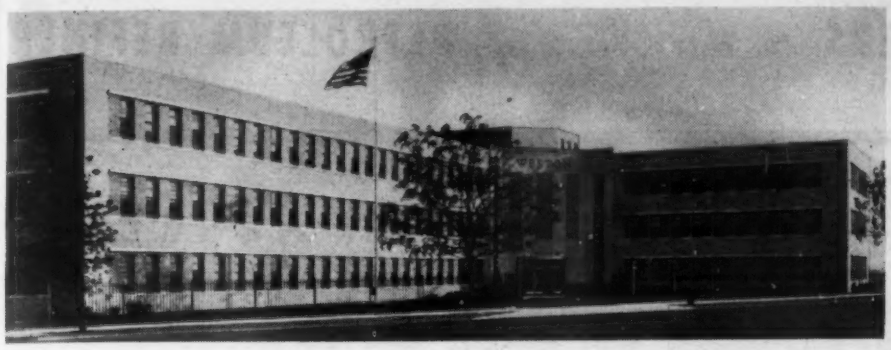
Jebens

New Agents Appointed

Leslie Company, Lyndhurst, N. J., has announced appointment of the following agents to handle industrial sales and service:

Seth A. Good & Company, Dallas, covering Northeast Texas; Equipment Sales Corporation, Kingsport, Tenn., covering East Tennessee; and, Fair Enterprises, Honolulu, Hawaii, covering Territory of Hawaii.

Weston Electric Occupies New Building



Weston Electrical Instrument Corporation is now occupying its new engineering and administration building (above) on the plant grounds at Newark, N. J. It is a three-story, T-shaped building of reinforced concrete with brick-faced hollow tile walls and a total floor space of 78,000 square feet. The new building permits expansion of engineering and administration facilities, and will release space required for manufacturing.

Air Reduction Changes

Harold H. Reed formerly assistant manager of Air Reduction Company's Metropolitan (New York) district has been appointed manager of this district, succeeding W. S. Schoenthaler, who is retiring after 37 years service. Reed has been with Air Reduction since 1923.

Oren M. Donohue, formerly assistant sales manager, has been appointed assistant manager, succeeding Reed and William B. Brower who joined the company in 1922 succeeds Donohue as assisitant sales manager.

Economy Buys Building

Economy Pumps, Inc., Hamilton, Ohio, has acquired from the War Assets Corporation the wartime addition to its plant, built in 1942 at a cost of $705,000. The new building, which contains 100,000 square feet, will house both of Economy's subsidiaries, Liberty Planers, Inc., and the Klipfel Manufacturing Company, which produces automatic regulating valves.

This extension of manufacturing facilities is expected to result in a large increase in the production of Economy pumps and Klipfel valves and furnish large flow-testing facilities for the Klipfel products, officials said.

Oronite Eastern Manager

Oronite Chemical Company of San Francisco, a Standard of California subsidiary, has appointed H. E. Bramston-Cook to the newly created position of general manager, Eastern operations. The executive has been general sales manager for the company for the past year.

Bramston-Cook will supervise both sales and product development work for Oronite east of the Rockies, and of foreign operations. His headquarters will be in New York.

ONE of the larger independent oil companies requires the services of several technically trained supervisors thoroughly familiar with catalytic cracking and also having experience with the operation of distillation units, thermal cracking, gas recovery and polymerization. Applicants should furnish complete outline of technical education, experience, previous employers, salaries received, and photograph. Address: 70-R, c/o Petroleum Refiner, Houston, Texas.

ONE of the larger independent oil companies requires the services of engineers capable of supervising refinery utilities, including generation of steam at high pressure, water treatment, water cooling and conditioning, and electrical and air distribution systems. Applicant should furnish outline of education, experience, previous employers, salaries received and photograph. Address: Box 60-R, c/o Petroleum Refiner, Houston, Texas.

Construction Engineers with experience in construction of petroleum refineries. State education, qualifications, experience, salary requirements, and date available. Address: Box 9-R, c/o Petroleum Refiner, Houston 1, Texas.

Construction Engineers with experience in construction of petroleum refineries and qualified to take responsible charge of large refinery construction projects. State education, qualifications, experience, salary requirements, and date available. Address: Box 8 R, c/o Petroleum Refiner, Houston 1, Texas.

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Alloy (Annealed)	Temperature, °F.	Tensile Strength, psi.	Elongation, % in 2 in.	Impact Resistance—ft/lb, Izod Specimens
Deoxidized Copper	Room	31,400	48	43
	—292	58,000	58	50
Cupro Nickel, 30%	Room	70,800	41	90
	—292	112,700	51	97
Cartridge Brass, 70%	Room	51,100	49	66
	—292	73,500	75	79

(Tests on 0.25 in. diameter specimens)

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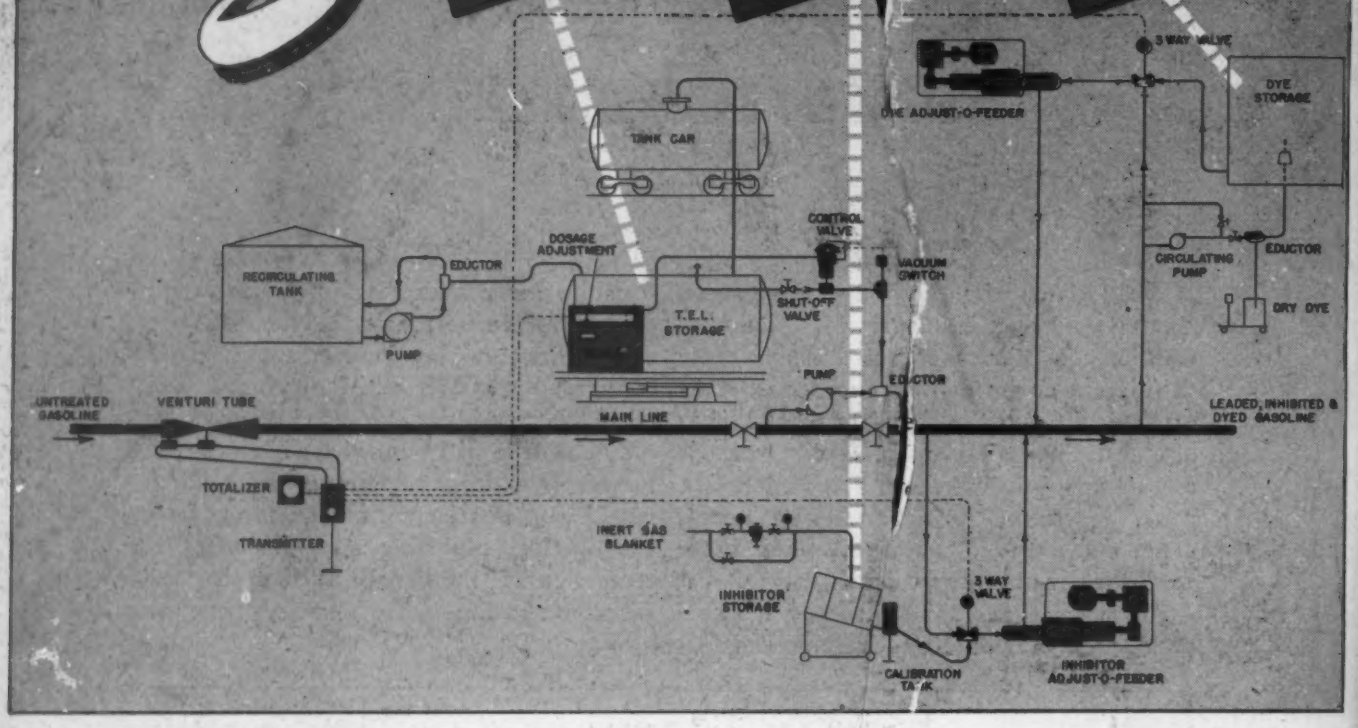
DYE



This schematic flowgram illustrates %Proportioneers% Treet-O-Control method of regulating lead, inhibitor and dye injection to a flowing stream of gasoline from a low loss Venturi differential producer.

Additives are handled in concentrated form and are taken directly from the containers in which they are shipped. Transfer of the lead fluid is made from tank car to weigh tank through a vacuum eductor system.

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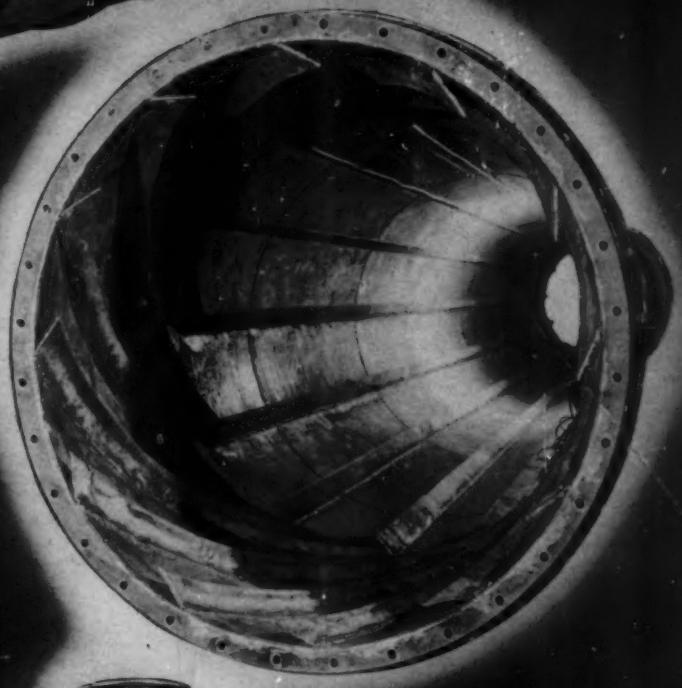

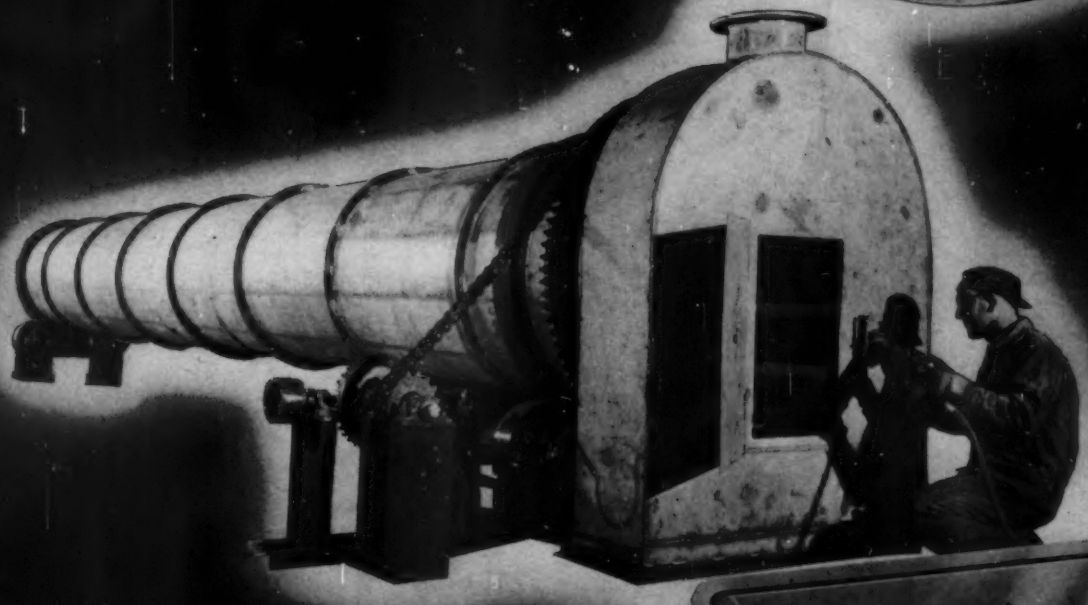

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